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REFINER
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The Look Box

C. I. O.?

SHOULD the C. I. O. continue its efforts toward unionization of the petroleum industry, the lines of battle will be more closely drawn with C.I.O. organizers, most of whom are new to the industry, on one side, and groups of company employees and the A. F. of L. on the other side.

Thousands of oil company employes are asking themselves the question: "What will the C.I.O. give me that I can not have in an organization of my own fellow company employes? What can they give me for my $2.00 per month that I would not have otherwise?"

A. F. of L. union members, of which thousands are employed at scale or better by the industry, are asking themselves the question: "What will happen to the ground gained through various A. F. of L. craft unions if the C.I.O. is successful in supplanting our unions with a mass union in which our immediate problems may in all likelihood be forgotten?"

And overshadowing all of it is the reputation of the C.I.O. movement for strife and disregard of law.

Public opinion is a powerful thing. Both the A. F. of L. and the oil company employes and their organizations recognize this. And now, after a few months of intensive activity on the part of the C.I.O., the C.I.O. movement has been rather closely identified with disregard of law and to the extent that it is getting in bad public odor.

To the complaint of the paid organizers that a company union is definitely under the thumb of the company the answer is that this is untrue and can not be true under the Wagner law. Even if it were true, and it is repeated it is not true, oil company employes are receiving and unquestionably will continue to receive fair treatment at the hands of their companies.

The new or current idea in the so-called company union or employes association is predicated on no company participation in its deliberations. The employes themselves in such associations as a rule elect their own representatives without the companies or the officials of the companies, in some instances even down to the foreman, participating. If they have grievances such grievances may be presented to representatives appointed by the company. Here, surely, is a method calculated to promote harmony.

No company appreciates being put in the position of having a paid agitator whose interest may be open to suspicion, to say the least, control negotiations for the company's employes. Likewise there is a feeling that company employes can better handle, present, clarify and secure equitable settlement of their own grievances when they deal with those problems themselves. In the case of associations of company employes, theoretically at least and practically in general, they are interested in the welfare of the company for which they are working. They are interested also in their own welfare. These two welfares are not in controversy but

have much in common. If some things arise which the employes as a whole feel are not fair, then they may be presented to the company's spokesman and the difficulties ironed out in a friendly manner. If the C.I.O. is successful men of the type of Lewis will be seen not simply representing the employes in any grievances they may have, but actually dictating the grievances.

Recent rather conclusive votes of employes of several companies seem to indicate that petroleum industry workers in the refineries and in the fields are not being misled by emissaries of strife who have picked the oil industry as a new battle field. The problem, however, remains one requiring solution by the individual petroleum industry worker.

Apples

THE apple is in the news again. This time it is no unnamed variety bouncing off the head of a Newton to call attention to the law of gravity, but a sample shipment of American apples which landed in British Guiana by airplane express 48 hours after the order was placed with the American exporter. The expeditious delivery enabled the pleased customer to cable an order for a larger shipment before the supply was exhausted.

Back of the news of the apple shipment; behind the news of ocean express via air, is the story of the work of petroleum technologists who have developed the motor fuels which make rapid air transport possible. The work was begun in 1928 when the U. S. Army Air Corps demanded a fuel which would permit development of greater horsepower without increase of motor size. Refinery technologists, scratching both pads and heads, proceeded to develop a powerful anti-knock aviation fuel of around 87 octane number.

Successful use of the new fuels in Air Corps planes indicated the production of about one third more horsepower per cubic inch displacement and substantial increases in flying speed, take-off, and climb, interested commercial aviation interests. The asked for standardized fuels in quality, with octane ratings between 70 and 87. Not only were the commercial fuels forthcoming, but they have been major contributing factors in the progress in speed and safety made by commercial airplane transportation during the past six years. Long distance flying was increased because the new fuels not only increased power output but decreased consumption. Hence the airplane express flying the oceans, and American apples on British Guiana breakfast tables.

The trend toward more effective aviation fuels is definite, and the U. S. Army Air Corps this year will use around a million gallons of 95- to 100-octane gasoline being produced in American refineries and by an increasing number of American refiners. Once more the refinery technologist is leading the engine builder, and is producing these new fuels in quantities more than sufficient for the motors that advantageously can use them.

New Crude

PETROLEUM from the Ten Section field and the Greeley field in Kern County, California, is wax bearing intermediate base oil, according to analysis of the Bureau of Mines. The oil is similar to oils of the same classification produced in the Mid-Continent fields, such as Seminole, Oklahoma and some of the fields in Kansas. The Section Ten crude has an A.P.I. gravity of 37.8 degrees, contains 0.37 percent sulfur, has a pour point of 10°F. and the color is greenish black. The approximate summary according to Bureau of Mines method shows 14.1 percent light gasoline of 75.1 A.P.I. gravity, while total gasoline and naphtha is 37.6 percent with A.P.I. gravity of 62.3 degrees. Kerosene distillate of 41.1 degrees A.P.I. gravity totals 4.3 percent. Gasoline is reported a 16.8 percent with A.P.I. gravity of 34.8 degrees. Non-viscous lubricating oil distillate totals 6.8 percent, medium lubricating distillate 4.5 percent and viscous lubricating distillate totals 3.6 percent. There remains 21.3 percent residuum of 11.7 degrees A.P.I. gravity. The carbon residue of the residuum is 12.4 percent and of the crude 2.6 percent.

The Greely crude has an A.P.I. gravity of 36.8 degrees, a sulfur content of 0.28 percent, pour point of 30°F. and the color is greenish black. Light gasoline of 73.3 degrees A.P.I. is reported at 8.8 percent, while total gasoline and naphtha is 33.2 percent with A.P.I. gravity of 57.9 degrees. Kerosene distillate of 43 degrees A.P.I. gravity totals 10.5 percent, and gas oil of 35.4 degrees A.P.I. is 15.1 percent. Non-viscous lubricating distillate totals 8.7 percent, medium is 6.8 percent and viscous is reported at .8 percent. The 22.7 percent residuum has an A.P.I. gravity of 14.7 degrees. The carbon residue of the residuum is 9.5 percent and of the crude, 2.2 percent.

The Bureau report states that the fractions heavier than gasoline and naphtha apparently are similar to corresponding fractions of oil from the Seminole pool of Oklahoma and other oils from Mid-Continent fields. The sulfur content of the crude is lower than that of most of the crude oil produced in California.

Handling Light Oils, Gasolines and Naphthas

L. C. TRESCOTT

POSSIBLY the products of the oil refinery giving least difficulty in handling are the naphthas, gasolines, the various burning oils. and the like. Persons outside the departments immediately concerned with these products are apt to consider that there are no troubles, that any equipment at hand and any layout will do the work. This is especially the view in many of the older and some of the smaller refineries. And while it is true that almost any equipment and layout can be made to work it is also true that some types of equipment and some methods of handling give better results than others. In any event it will be of some profit to look into the various problems that come into this end of refinery operation.

Oils of this grade handled in large quantities in the average refinery may be roughly classified as products going into motor gasolines or those going into burning oils. Special cuts may be made for various purposes but these may be handled in the same way as the products mentioned above.

In the days of shell stills it was necessary for the refined-oil department to conduct extensive distillation operations of its own, particularly where much crude was run to coke and a rather broad naphtha cut was taken off the primary stills. With the use of continuous-column stills this is unnecessary as it usually is possible to cut naphthas and light oils to specification on the primary distillation. Modern pressure stills also produce a finished product so far as distillation is concerned.

SPECIAL CUTS

Special cuts of light oils are made either in the distillation or in the refined-oil department. Some of these may be taken direct from the continuous stills but it is not always easy to take a narrow cut off a continuous stream unless the still has been designed for that particular job. If any variety of such cuts is to be made it is well to provide a batch still for the purpose. A cut can then be taken on the large still which will carry a high percentage of the desired stock. This can be charged to the batch still and brought to specification as a bottom rather than as an overhead product. This procedure is especially recommended where the cut desired is of close specification within a narrow distillation range and where its volume is a low percentage of the continuous stream from the large still.

The first move away from the stills is to the running tanks. In the final delivery of the oil to trucks or cars high accuracy is necessary in measuring the oil loaded and close gauging must be done to check shipments and to keep losses down, but in the refinery the condition is a little different. Most of the time it is necessary to pump running tanks with the stills running into them. Records will show more oil received than the gauges of the tanks pumped will indicate. It is also possible that the treaters or other equipment may be drawing from the receiving tanks while a pumping is coming into them.

MORE TANKS

The experts will point to a very obvious way around this difficulty—provide an extra set of tanks. Unfortunately we must deal with things as they are —and whoever heard of a refinery which really had enough running tanks, or the room to place enough running tanks, especially after it has been in operation a few years. Of course it is possible to make arrangements to receive the stock in tanks that are not moving and usually this is done, but in any event figures of sufficient accuracy for yield purposes or for refinery transfer purposes can always be dug out somewhere along the line.

As absolute accuracy in gauging is neither necessary nor possible floating roofs can be used on small tanks. If not some other vapor saving device should be installed. These are especially valuable on still running tanks for naphthas as they are constantly in motion and the temperature often runs higher than that of the outer air. Particularly at seasons of the year when the temperature of the cooling water is up the products from the stills are likely to run quite warm.

Stock tanks should be large and operated under such conditions that they may be gauged accurately. It may be well to note a difference in gauging and handling conditions in the refinery and in those of the bulk or delivery station. In the refinery the pumpings from stock tanks are fairly large, and are generally made from one large stock tank to another. There is a more or less continuous stream of pumpings of the same products. Any gains or losses tend to compensate each other, at least within the limits of accuracy required by refinery processes.

TYPE OF TANKS

In the bulk station or on the refinery loading racks the single deliveries are small, most of them of such a volume as to make little difference in the level of a large tank. There will often be an appreciable difference shown between oil loaded through a meter or into a gauged tank truck and that indicated by a large

tank. Customers keep a close eye on the oil they receive and it is convenient to be able to fall back on tank gauges in case of controversy. In order that all figures may be kept accurate the tanks delivering to the loading racks are often of a tall, narrow type, in which a small quantity may be gauged accurately and at the same time large enough that filling them will give an accurate check with the stock tank.

Final deliveries from the loading rack are measured either through meters or by the calibration of the tank cars or trucks. In the case of large trucks hauling regularly from the rack the compartments are loaded to a "finger" which is set at a determined capacity. Usually the volumes and the finger settings are checked by the sales department before such loadings are authorized. Smaller tank loadings, less than compartments, or trucks which have not been calibrated are loaded through meters. These should be checked frequently. This last may seem too obvious to stress here, but the fact remains that an astonishing amount of complaint does come back to the loading rack due to the fact that meters need adjusting.

In laying out the lines to carry naphthas and burning oils every precaution must be taken to prevent contamination, particularly to prevent color loss. The various grades of naphtha may be pumped over the same set of lines as may the burning oils but no connections or jumpovers should be permitted between these lines and those carrying heavier or darker oils. In clearing the lines after pumping no air can be used to blow them as such practice is unsafe. In the few cases where displacement is necessary water may be used.

SELECTION OF PUMPS

The pumping of naphthas and light oils is generally not considered a problem. Any pump that is sufficiently tight may be found on this service and usually functions at least well enough that there is no great cause for complaint. As in everything else, however, there are some methods which give better results than others. There are points of installation which cut down efficiency. There are types of equipment which will make operation easier and at the same time will cut down the cost of moving the stock.

The handling of light oils is a field for the centrifugal pump. Many oil men will not agree with this and will install nothing but reciprocating pumps for any purpose. It is true that they have probably had unfortunate experiences with centrifugals and it is also true that the reciprocating pump is flexible, that it will work on any kind of oil and almost under any kind of conditions. It can be slung into any opening in a pipe line and will move the oil in that line. The centrifugal pump must be selected and installed for the particular service which it is to perform. Neglect of this point is the direct cause of any dissatisfaction with centrifugal pumps. However there are many places particularly where light oils are to be handled, where a direct-connected electrically-driven centrifugal, or a turbine-driven centrifugal will show decided economy over a reciprocating steam pump, and it is worthwhile to so select and install the centrifugal that these economies will be realized.

Trouble generally comes to the centrifugal pump through the suction end. This is equally true of a reciprocating pump but the latter will pound along at a lowered capacity under conditions that will stop a centrifugal entirely. This would seem to be something of an argument against the use of centrifugals. It is, as a matter of fact, the argument which has kept them out of more general use in the petroleum industry, and if we are to refute it we must get busy on the suction end.

With any oil or other liquid, air is likely to trap in the head or the casing of the pump. This applies to either type. With the centrifugal it means that the pump will not move oil. In more cases than not this trouble comes primarily from the suction line itself. No pump handling naphthas should have a suction head on it. This means that the suction lines should be as large as can be fitted to the pump and that the pump should be placed as close as possible to the tank or the source of liquid which it is to move. There should always be a gravity head on the suction side of the pump. A practice which works out very well is to use a centrifugal pump direct connected to its motor and placed right alongside the tank or tanks for which it is to work. This may be housed in a small weather-proof box and controlled by a push button located in the pump house, the loading rack or any desired point. All old stuff, it may be said. Right enough. But take a good look around the plant. How many pumps are operating through extra long suction lines? How many centrifugals have been stuck in where a reciprocating pump originally was installed? How many different sizes of pipe are in those suction lines? It is a safe bet that plenty of them start out large enough, say a six-inch. Then they jump to a three, followed possibly by an eight, with maybe a short two-inch section between it and the final six-inch that hooks to the pump. It is done every day and the older the refinery the more such lines are found. And when found in the suction line on a naphtha pump there is no use in expecting the pumps not to get air-bound or gas-bound, particularly in warm weather.

Of course it is not always possible to pick an ideal location for the pump. Many times pumps must be installed on existing lines and it is not possible to improve conditions on the suction end. Here two things will help. Slow down the pump until it fills on every stroke so that it is operating under such a head that gas does not separate. If it be a centrifugal close down on the discharge valve until the pressure is sufficient to keep the gas from separating. In this connection it must be pointed out that it is possible to run a pump, even a reciprocating pump, too fast, especially when pumping naphthas. The pump should never travel at such a rate that a real suction head develops. The extra speed will never make up for the capacity lost due to the gas separating and preventing the pump chambers from filling to their full extent. Further, the extra speed is hard on the pump.

For pumping gasolines and light naphthas the specially designed vertical discharge type of centrifugal pump should prove of value. There are several of this type on the market which have been developed for pumping acids, hot liquids and light products in the chemical industry. Trapping in a centrifugal occurs at the very top of the casing. Even when the outlet of the usual tangential discharge type of centrifugal pump rises vertically, there is opportunity for gas-binding as in such an arrangement a part of the casing is above the point of discharge. The type of pump mentioned above may either have the main discharge on a vertical tangent with an auxiliary opening at the top of the casing to feed any gas that may collect into the outgoing stream. Or, the

discharge of the pump may be in the form of a hopper at the top of the casing. The general idea is to keep any gas or vapor in the moving stream. This type of pump will work under a greater suction head than the usual type of centrifugal.

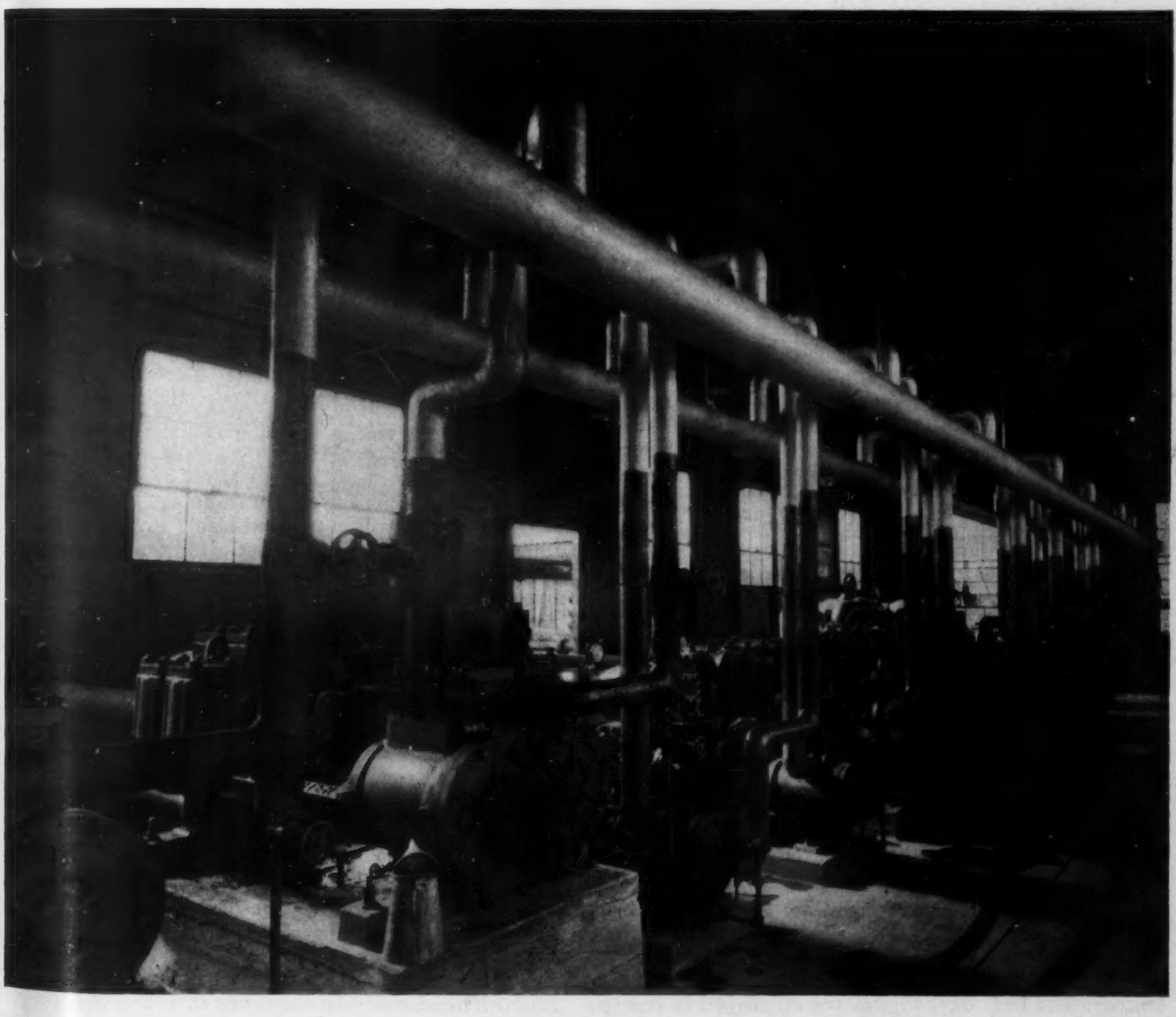
In the ordinary type of centrifugal pump it is possible to relieve a gassing condition by taking a pipe from the top of the casing and carrving it into the discharge line of the pump. This will carry off any separated gas and keep the casing always full of liquid, though in many cases it is not possible to put in a pipe large enough to get rid of the trouble entirely. If such a condition is expected it is better to install one of the centrifugals with the specially designed vertical discharge.

Chambers may be built which will keep a centrifugal pump always primed. If there be a suction head on the pump, either due to a long suction line or to the pump being higher than the source of supply, such a chamber will improve operation.

With the increased use of light solvents in the refinery the pumping problems are bound to become more complicated, and here especially if a centrifugal pump is to be used it must be suited to the job. While it is often claimed that a centrifugal pump will not "suck," as a matter of fact there have been developed for hydrometallurgical and other purposes pumps which will pull from under as high as 25-27 inches of vacuum. These pumps are of a simple single stage type and are provided with specially designed sleeve bearings which prevent a leakage of air into the pump. These have been in successful use for years, but they are for suction purposes only and usually are installed to discharge at atmospheric pressure.

In such cases as where a clean oil mixture, free from abrasive clay or filter medium, is being handled there is no reason why a reciprocating pump cannot be used. This will pull from under the vacuum and discharge at any desired pressure.

In the foregoing it may seem that the writer has stressed points which are obvious. The fact remains that in the petroleum industry a very large number of centrifugal pumps have been installed without being given proper consideration for the work they are to do. And more failures than not are caused by the suction line faults shown here. With the increased electrification of the refinery more centrifugal pumps will come into use and if the full efficiency of the electrification is to be realized it is necessary that these pumps be fitted to their job and that they be properly installed.



Transfer Pump Room

Design and Operation Features of

Black Gold Cracking Installation

J. C. ALBRIGHT

BLACK Gold Refining Company, Oklahoma City, like many other refining companies that have come into prominence during the past few years, has grown from humble beginnings. Originally built as a simple skimming plant to process the oil produced in the discovery section of the old Oklahoma City field, new equipment has been added from time to time until sufficient equipment is available to permit manufacture of almost any fraction from steam-refined cylinder stocks to high-octane cracked gasoline. The latest addition to this refinery is a complete Dubbs cracking unit, employing two furnaces for selective cracking, so that the topped crude from the skimming plant, and some of the lighter fractions may be converted to motor fuel.

Crude is pumped to the primary reducing unit where straight-run gasoline, naphthas, tractor fuel, kerosene and steam refined cylinder stocks are manufactured, with the surplus topped crude charged directly to the cracking unit.

Reduced crude from the base of the flash and fractionating columns is either passed through a continuous system for the manufacture of steam-refined cylinder stocks, or, is used as charging material for the Dubbs cracking unit. Steam-refined cylinder stocks made in the continuous battery of reducing equipment is passed through coolers directly to tank car, or to storage.

When the wells in the south portion of the field were new, and flush production was being taken from them, salt water did not present any sort of a problem for the refining of this type of crude. Since so many of the older wells are now almost in the stripper class, or are being produced with the water table materially raised, salt water is being produced in steadily rising quantities.

Due to the absence of salt, and the characteristics of the oil, relative to sweetness, corrosion in refineries was not severe several years ago and no provisions were built into plants for the removal of hydrogen sulfide and hydrochloric acid generated from various substances responsible for this type of difficulty. Neither was trouble encountered by depositions of salt and other chemicals within the tubes of heaters and condensers, towers and other accessories. At present, however, salt must be removed, as well as hydrogen sulfide, somewhere in the stream, either before the oil enters the tube heaters, or before the gasoline is finished.

The salt content, particularly, is not large, compared with oil produced in other sections, but what is brought in with the oil, crystalline or in suspension, is removed by settling, prior to passing the charge through the skimming section so that chlorides will not attack condenser sections and vapor lines. Low-temperature heat exchangers are employed in removing the salt from the oil, together with a settling drum having a capacity sufficient to cause the salt to drop out before the oil enters the portions of the plant where higher temperatures are employed. No water mixing or chemicals are used to cause the separation of salt. Application of moderate temperatures are sufficient to cause adequate separation while the charge is slowly passing through the drum. The flow is through the vessel in the usual manner; in at the top of the shell at one end, and out at the other, with a water drain placed at the lowest point which provides for constant or intermittent bleeding.

A tubular preheater permits removal of a portion of the more volatile fractions from the crude charge before the stream of oil is passed through the regular pipe still. This unit is of the upshot fired type, with the stream of oil flowing counter current to the gases of combustion. As the temperatures used in this heater are in the neighborhood of 350° F., large thin-walled tubes are installed. Increased production of gasoline and other products is secured and this addition provides greater flexibility of the plant as a whole.

Crude is processed through the skimming pipe still which has a transfer line leading to the usual type of fractionating column and flash tower commonly found in plants producing straight-run products directly from the crude. Naphthas, straight-run gasoline, kerosine, tractor fuels and topped crude are the products ordinarily made in the primary distillation of Oklahoma City crude, with each fraction kept separate until such time that requirements justify blending. Bottoms are removed from the base of the columns by pumping to storage, when used for cracking, or, pumped to a continuous battery of shell stills for the manufacture of steam-refined cylinder stocks, which are finished in the unit with comparatively large quantities of steam, and are in condition for shipment to markets as they emerge from stills.

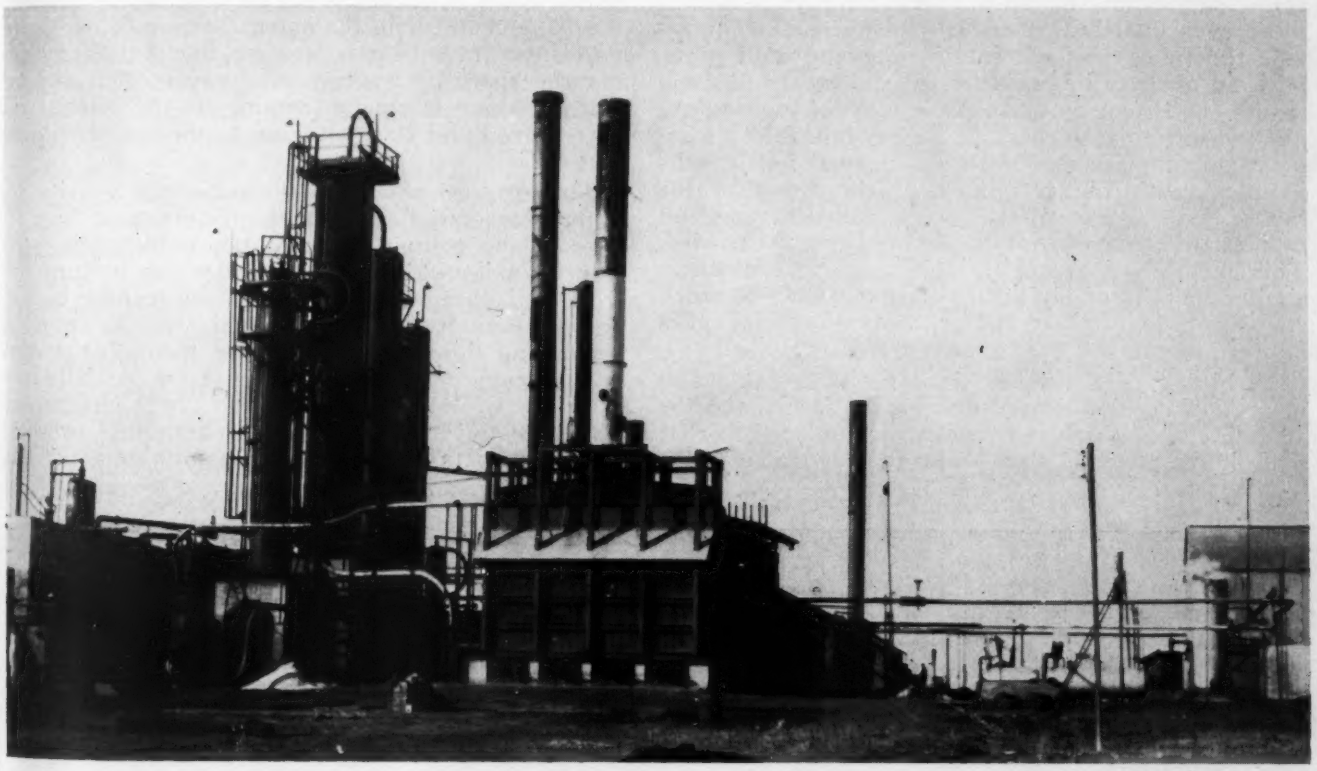
The topped crude which is designated for cracking is picked up by a small duplex steam operated piston pump at a rate of approximately 750 barrels per day. and charged to the flash column of the carcking unit where the processing of the stock is begun. This column, the base of which is used for the withdrawal of fuel oil, has 6 side-to-side plates installed in the upper section below the nipple where the topped crude first

enters. Above this section, bubble plates are installed to maintain fractionation. The bottom section, in which the fuel oil accumulates from the line leading from the reaction chamber, is without baffling and is fitted only with a weighted spray for washing the walls with the material from the reaction column.

The pressure maintained upon this column is about 62 pounds, gauge, with a top temperature of about 740° F. Topped crude charged to this column is stripped of a large portion of its lighter fractions. Heavier fractions that will not pass overhead with the stream of vapors from the cracking unit reaction chamber stream drop back through the side-to-side panel and are accumulated in a draw-off section, which has no down-comer, but is equipped with a stack, so that vapors may pass without bubbling through the liquid trapped in this section. The liquid is removed from the tower by the heavy oil hot-oil pump and is passed to the heater for processing in the heavy-oil heater.

The section of the flash column at this point is equipped with an internal float having an arrow, or pointer attached to a shaft extending through a packed bearing to indicate to the level of the fluid in the section. The indicator is also connected to a remote controlling instrument which regulates the speed of the topped crude charging pump. Air tubing leads from the remote controlling valve on the pointer shaft to the back pressure diaphragm valve on the discharge of the topped crude charging pump. This valve, also connected by tubing to a second diaphragm valve, controlls the steam furnished as power to operate the pump. When the pressure exceeds that required to move a definite volume of topped crude to the flash tower to maintain the required level in the accumulator section, the valve, controlled by the instrument on the end of the indicator arm, causes the valve in the stream line to close a proportionate amount, so a piston pump may be controlled directly by a valve in the discharge line, and indirectly controlled by a valve in the steam line leading to the steam chest of the unit.

A suction line is attached to the lower part of this section which leads to the heavy oil hot-oil pump, a unit having duplex outside packed pistons 4½ by 12-inch, and operated by a 12 by 12-inch horizontal single cylinder steam engine with a self aligning coupling between the engine and the pump. With normal throughput, this pump is operated at a speed of 2880 revolutions per hour and with a discharge pressure of approximately 475 pounds, gauge. The topped crude which enters the flash column above the side-to-side pans has a normal gravity of about 31° A. P. I., and after becoming heated is ready to pass through the heavy oil cracking furnace.



Dubbs cracking unit, Black Gold Refining Company at Oklahoma City, Oklahoma.

This furnace is one of the first to be constructed of this particular type and design. The heater contains 72 tubes, part of which are in the convection section and a part in the radiant banks. Eight burners are installed on each side of the furnace, with controls on the outside of the walls. One fuel riser is placed for each pair of burners, but each burner is equipped with an individual lubricated plug valve for auxiliary control of gas. Tuyere fittings are installed just inside the outer refractory walls of the furnace with the adjustment made so that the flames impinge against the refractories rather than against the tubes and hangers in the center of the combustion chamber. Gas to the burners is controlled by valves which feed the required volume to the burners. Dampers to control the mixing of air and gas to the burners are placed outside the burner setting and may be varied to suit combustion conditions.

The arrangement of the cracking coils within the

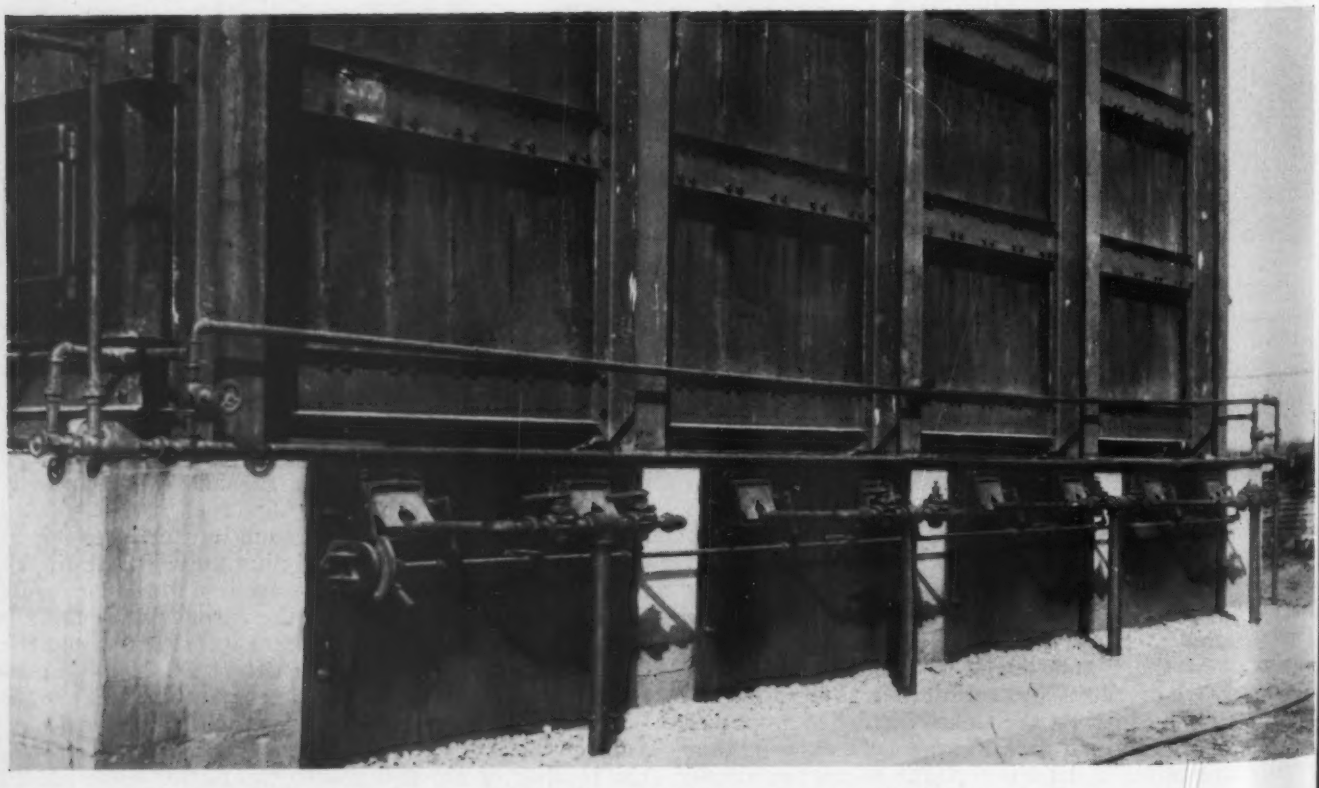
furnace include 4 rows of 8 tubes each, made of 4/5 chrome, ½ percent molybdenum alloy, and followed by two rows of 4 tubes each, which constitute the convection section of the furnace. The radiant section of the coils is made up of two rows of tubes, stacked upon hangers, one above the other, each row containing 16 tubes. The top 4 rows of tubes of 8 tubes each have an inside diameter of 2⅜ inches, and the following 2 rows, containing 4 tubes each, are 2⅛ inches inside diameter. The vertical bank of tubes are also 2⅛ inches inside diameter, but made of high carbon steel. Headers were placed on the various banks and enclosed within a sheet metal housing which protects them from the elements.

The arrangement of the vertical radiant banks of tubes is such that the tubes are staggered on the hangers, so that the heat radiated from the hot walls and the burners on each side of the furnace will not be shadowed, or shielded from any part of the exterior surface of the tubes, permitting, it is said, a more even distribution of heat to the oil flowing inside the tubes, and eliminating hot and cold spots, with an absence of buckling. All tubes are uniform lengths of 24 feet, so that shop lengths of tubes might be used for replacements, should any fail. Inlet, crossover, and transfer sections are constructed of lengths identical with the remainder of the tubes in the heater, with additional features of rolled flanges and nipples to compensate for additional lengths needed. This method of arranging the tubes and utilizing similar lengths of piping makes it possible to stock standard tubes without the necessity of special tubes for certain parts of the furnace.

Turbining of the tubes during a shut-down and inspection periods is arranged for by special ladders at each end of the furnace with rungs placed so that scaffolding may be placed at any convenient point for workmen to stand while handling the hose. The stack is relatively short and insulated a short distance from the breeching to protect it from deterioration due to hot gases from the furnace. The walls of the furnace are of the full suspended construction, with refractories placed on the interior, and insulated on the exterior surface with rock-wool. The flow of oil through the furnace tubes is from top to bottom contra to the flow of gases from the tuyere burners, which is upward.

The charge flowing from the cracking tubes in the furnace through the transfer line enters the reaction chamber at the top and at the same point that the light oil furnace discharges its material, and both streams, mingled together, follow the usual course from top to bottom. No liquid is held in the reaction chamber, except that amount which is sufficient to break the pressure down from the standard of about 270 pounds to that maintained on the flash chamber, which is about 62 pounds. The fluid and gases enter the flash chamber in the empty section below the removal, or draw-off pan, and are distributed throughout the space by a weighted spray so that any formation which might be forming on the interior will be removed and washed down to the outlet with the fuel oil.

The stripped residuum controlled for gravity and other characteristics is constantly removed from the base of the column through the bottle-neck extension by a liquid-level controller which in turn controls a diaphragm valve on the line leading to storage. Vapors from the combined streams from the heavy and light oil furnaces rise through the stack in the heavy oil draw-off pan, and are partially fractionated, while fractionating the crude charge, and leave the tower through the conventional overhead vapor connection. Led through a suitable vapor line,



Firing side of heavy oil furnace. Each burner is equipped with an individual lubricated plug valve for control of gas.

they enter the bubble tower near the base for final treatment.

The bubble tower is of conventional type, being fitted with bubble plates which control the overhead stream, together with the liquid reflux pumped back over the column, so that the final product will not exceed an end-point of about 375/380° F. The top of the column is held at about 350° F. Condensation of the overhead vapors is accomplished by passing them through conventional submerged coils.

The bubble column is fitted with a liquid basin in the lower end which catches the heavy material knocked back by the action of the bubble trays and the pumped reflux, to be utilized as the material processed through the light oil cracking furnace. A large steam-driven piston pump handles this material and takes suction directly from the bubble tower and delivers the oil under a temperature of about 570° F. to the light oil cracking furnace at a pressure of 400 pounds.

While processing this class of material, the convection outlet temperature averages about 735° F., while the transfer-line temperature runs about 975° F. The pressure on the outlet of the furnace is controlled by the pressure of the reaction chamber, which is 270 pounds during normal operation.

Vapors from the cracked gasoline accumulator drum are transferred to an absorbed, operated at about 20 pounds, gauge, which takes material from the outlet of the bubble tower for absorption oil, cooled before using, and returned to the cracking plant with the absorbed gasoline vapors, and is not used again, as such, but redistilled and blended back with the next cut for the light oil charging pump.

The pressure distillate, or cracked gasoline is handled through a low-pressure fractionator, or stabilizer, operating at 21 pounds, and the reboiler carried at 150° F. Vapor pressure of the product is varied according to the seasons, but normally runs from 8½ to 9 pounds, Reid. Vapors from the stabilizer are passed through an absorption column before being burned in the furnaces, to strip what low-boiling fractions remaining in them, such as part of the propane, iso-butane and the normal butane necessary to discard to hold the vapor pressure of the pressure distillate to the desired point. Third-grade gasoline is used normally for this agent, so the initial boiling point and other characteristics may be changed without the necessity of adding a greater part of natural gasoline for the purpose.

Finished cracked gasoline, as it comes from the stabilizer, is pumped through treating equipment to remove the objectionable sulfur compounds with doctor solution. Prior to contacting with doctor, the gasoline is given a caustic wash to eliminate the hydrogen sulfide. Treated gasoline is run from the treating tank settlers to storage.

Water for plant purposes is produced on the premises by two wells and a rainfall catch basin. One of the wells, which is relatively shallow, produces water with a considerable amount of permanent solids in it, while the deeper well water is soft, and is suitable for boiler feed as it comes from the tubing without softening or treating for permanent hardness. Excess soft water is used to replenish the cooling system water, with the over-flow from the shell and tube coolers, the submerged coolers and other unit passing to the catch-basin, where it is picked up and used again.

The control room of the cracking plant is divided and a part of the building houses the heavy oil hot-oil pump and the topped crude charge pumps. Glass walls allow the operators to watch through windows the operation of the plant and instruments on the columns, without the necessity of leaving the principal controls at the instrument board. Indicating instruments attached to the various thermocouples in the plant are mounted upon the control board, as well as the recording thermometers, flow meters and control units. Firing of the heaters may be increased or lowered by merely manipulating a small control at the lower left hand corner of the control board, so, with the plant in good condition, it is seldom necessary for the operator to leave the control room.

Main column in cracking plant of Black Gold Refining Company, showing float indicator above manhole which controls the suction of the heavy oil hot oil pump.

De-Salting Unit

Operation at

Michigan Refinery

PROCESSING crude from the fields in central Michigan, particularly that from the Crystal, Porter and Midland districts, Naph-Sol Refining Company has encountered a difficulty with salt produced in large quantities with the oil.

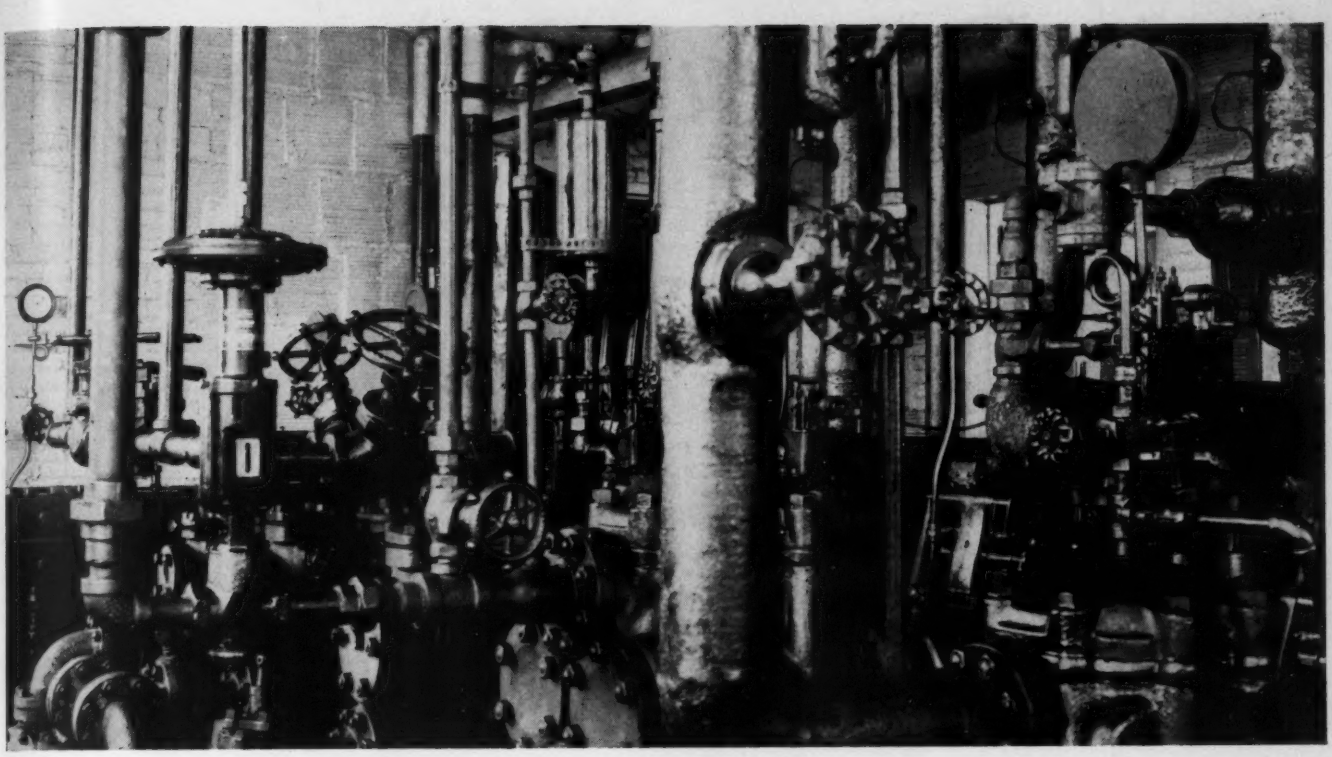
Producers in most of the fields are using field equipment to remove the emulsions and salt water that can easily be separated but there remains about 150 pounds of salt per 1000 barrels of oil, which must be removed at the refinery. Naturally, when oil having as high a concentration of salt as this is processed corrosion takes place, as well as still tube plugging.

By mixing fresh water with the oil after it has been run through the skimming plant, and heating it by exchange with residuum from the cracking plant, settling is accelerated and the salt content is reduced to a point where satisfactory operation is obtained. The method used by Naph-Sol Refining Company includes a fin-type exchanger combined with a larger, horizontal separation chamber, equipped with 250 feet of 4-inch tubes, through which steam is passed to assist the heating of oil at times when exchanger service is not sufficient, but particularly when the unit is being brought on-stream after a clean-out.

The operation of the unit begins at the topped crude charging pump, which is fitted with a water line connected to the intake manifold so that the fresh water from plant storage may be injected. The volume used is controlled manually, amounting to about 4 percent of the total charge to the cracking plant, but varies with concentration of salt in the particular crude being processed. Tests are conducted upon each batch of oil after it has been skimmed, and water volumes determined accordingly. The bulk of the salt in the topped crude is sodium chloride, with a small amount of magnesium chloride. These are reduced to about 20 pounds per 1000 barrels of charging stock as the oil is passed through the desalting unit.

Partial mixing of the fresh water and the cold

Desalting drum installed in the Dubbs Cracking Plant of Naph-Sol Refining Company, Muskegon, Michigan



Water connection to topped crude charge pump for de-salting crude charge.

topped crude is obtained in the fluid end of the reciprocating pump, assisted by the turbulence in the line from the pump room. The oil and fresh water mixture travels first to a heat exchanger unit flowing countercurrent to the residuum from the cracking unit flash chamber to raise the temperature of the oil and water to about 300° F. Additional intimacy is assured as the oil and water passes through this unit due to the construction which causes increased turbulence of the fluids.

The second unit in the desalting apparatus consists of a horizontal chamber, 5 feet in diameter and 42 feet long, covered with 3 inches of rock wool which is protected from the elements by a weather proofed asbestos outer coat, to maintain the applied temperature on the oil and water mixture so that the salt will dissolve more rapidly and settle readily from the charge with the water. The drum was installed with the water draw-off end about 2 feet lower than the oil outlet, to provide as small a space as possible for water accumulation with respect to the amount of fluid passing through the unit.

Since the drum is installed with one end lower than the other, the larger part of the brine is withdrawn from the lower end through a manually controlled bleeder valve. A second draw-off connection is attached to the drum at a point about midway of the lower side of the shell so that the interface between the oil and water can be determined. The draw-off line at the lower end of the drum always removes clear water, and, while the drum is in operation, the upper connection, bleeding a very small amount of water, occasionally bleeds dark fluid, which shows that the oil and water separation line is at the desired point in the drum. It is the intention of the operators of the cracking unit to set the upper draw-off bleeder valve so that water will run consistently clear, but may be changed to the darker shade quickly. It is desired to maintain a water level just slightly above 1 foot at the lower end of the horizontal drum.

When bringing the cracking unit on stream, with all units cold, the 250 lineal feet of 4-inch tubes within the horizontal drum are used, with steam from the boiler plant passing through and governed by a control valve on the inlet line. Outside heating is useful to remove the salt from circulation to prevent deposition within the exchangers and upon the stripping plates of the fractionating column. When the circulation has obtained sufficient heat from the oil after passing through the tubes of the heater, the steam is cut from the internal coils and the desalting unit operated with exchanger heat.

As tests are conducted upon each tank of reduced crude to determine the amount of salt contained in it, the water volume is adjusted by a hand control valve at the intake of the topped crude charge pump. The capacity of the horizontal drum is equal to 1½ hours charging, constructed with this size to provide adequate separating time. Consequently, the salt must run considerably higher than any average crude to cause pulling over with the oil of any water which may be introduced into the stream at the charge pump. In order to prevent ebullition of the fluid under the temperature of about 300° F. a back pressure is maintained upon the stream of topped crude, averaging about 170 pounds, which, it is said, provides sufficient capacity for a considerable overage of fresh water.

The cost of this unit was around $2000 installed which cost, when compared to the amount of topped crude refined each day, is not excessive, particularly when the plant was shut down once every 15 days for a clean-out before the unit was installed. Running time has been about doubled, and corrosion reduced considerably. The operating cost cannot be considered as an additional expense, except for an occasional piece of repairing, as it utilizes heat obtained from the residuum of the cracking unit, and fresh water from the company wells.

Salt Removal from Crude Oil . . .

Chemical and Physical Methods

E. R. JONES
Skelly Oil Company

THIS paper was not prepared for the purpose of presenting a method that would serve as a cure all for all problems dealing with the desalting of crude oil. Neither does it present the results of years of deliberate research on the problem. Instead, the paper will discuss the observations made and the results obtained while trying to find an immediate practical solution of the problem.

Throughout the discussion, the word "salt" does not refer to sodium chloride alone, but to the chloride content whatever the positive radical may be. Whenever quantities are mentioned, the results are expressed as sodium chloride.

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W P R A

THE first question, of course, is why need the salt be removed from the crude? The first answer is the corrosion trouble that is traced directly back to the salt. This problem has seemingly increased with the increased acidization of wells. The crude that is obtained from a dolomitic limestone will contain varying amounts of magnesium chloride if the producing well has been acidized with hydrochloric acid. The purpose of acidizing is to break down or make more porous, the limestone structure. The acid reacts with the limestone to form carbon dioxide gas, water, and the chloride of the positive radical that thad been combined with the carbonate whether it had been calcium or magnesium. If magnesium chloride is present in the crude it will, by hydrolysis, revert back to hydrochloric acid which is very corrosive to iron. This type of corrosion is first noticed in the first stages of distillation and is combated by the use of ammonia, a relatively expensive chemical. This corrosion is noticed mainly in the condensing section of the units.

Sodium chloride and similar salts are suspended in the crude. They deposit on the heater tubes and still bottoms causing hot spots. It is also claimed[1] that they act catalytically to produce increased amounts of hydrogen sulphide.

PLUGGING OF EXCHANGER TUBES

A second reason for salt removal is the plugging of exchanger tubes and condenser sections because of the presence of salt in the crude being processed. The salt usually is present in a water solution emulsified with the oil. If the crude is passed through preheating exchangers, and sufficient temperature is reached in the exchanger to vaporize the water, salt will adhere to the tube walls, eventually plugging the tubes and necessitating a shut-down for cleaning. If the hydrochloric acid formed during distillation is neutralized with ammonia, the condensers and vapor exchangers, and indeed even the trays in the fractionating tower, have in some instances become plugged as a result of the deposition of ammonium chloride. Careful pH control tends, of course, to hold down such deposition to a minimum.

The third probable answer is the corrosion of still tubes. This is a question that deserves consideration and study in the future. There is more than surmise

to this question. The corrosion seems to be confined to the tubes handling the heavier gas oils. Lime addition has been found greatly to reduce this corrosion. One refiner has installed a rather efficient desalting unit and no difference is seen in the rate of corrosion in cracking still tubes. However, less coke deposition is being obtained in the topped crude viscosity breaking coil. This action is attributed to the removal of salt from the crude.

The fourth answer to the refiners that are making road oils and asphalt is the need for meeting such specifications as solubility and organic content on those products. Salt in the residue often brings about rejection of these products.

Salt exists in the crude in a water solution emulsified with the crude and also in crystalline form probably surrounded by wax crystals or perhaps by amorphous wax. Because of the method of producing crude oil, the oil and brine are intimately mixed and it is practically impossible to prevent the formation of emulsion.

In a study of this problem, crude oil taken immediately from a well was shaken with ten percent water, allowed to stand for one hour, and was then analyzed for salt. Seventy percent of the salt was removed in this manner. This suggests the possibility that the oil-brine emulsion tends to become stabilized with age. To remove the salt at this point is a much easier job but it means treating each well, which is not ordinarily practical. The refiner has to deal with the emulsion after it has become stabilized and that is what will be discussed in this paper. The subject, "Salt removal from crude oil — chemical process," is possibly not quite accurate for the method is more physical than chemical.

The salt content of the crude can be reduced by allowing long-time standing of the crude. The following are data taken on a crude tank that had been allowed to stand four days. The samples were taken from the charge pump at the stills.

Time	Salt—Grains/Gal.
9:30 A.M. 4-13-35	24.9
11:00 P.M. 4-13-35	25.2
11:00 P.M. 4-14-35	16.5
11:00 P.M. 4-15-35	18.5

The crude at the top of the tank had 70 percent as much salt as the crude at the bottom. Not all refiners have sufficient crude storage capacity for such a treatment. Even so, there will be a pay day some day on the tank bottoms.

The salt content of the crude varies with the season as is shown by Figure I. There is less salt during the warm summer months than during the cold winter months. This suggests the idea of lowering the viscosity of the oil with heat and thus aiding to separate the brine from the crude.

BREAKING OIL FILM ON BRINE

If the salt is present in an emulsion of brine in oil, the small particles of brine are dispersed throughout the oil phase. The very minute particles of water are unable to coalesce because of a resistant film surrounding the watre particles. Whenever that film is broken, the small brine globules will con-

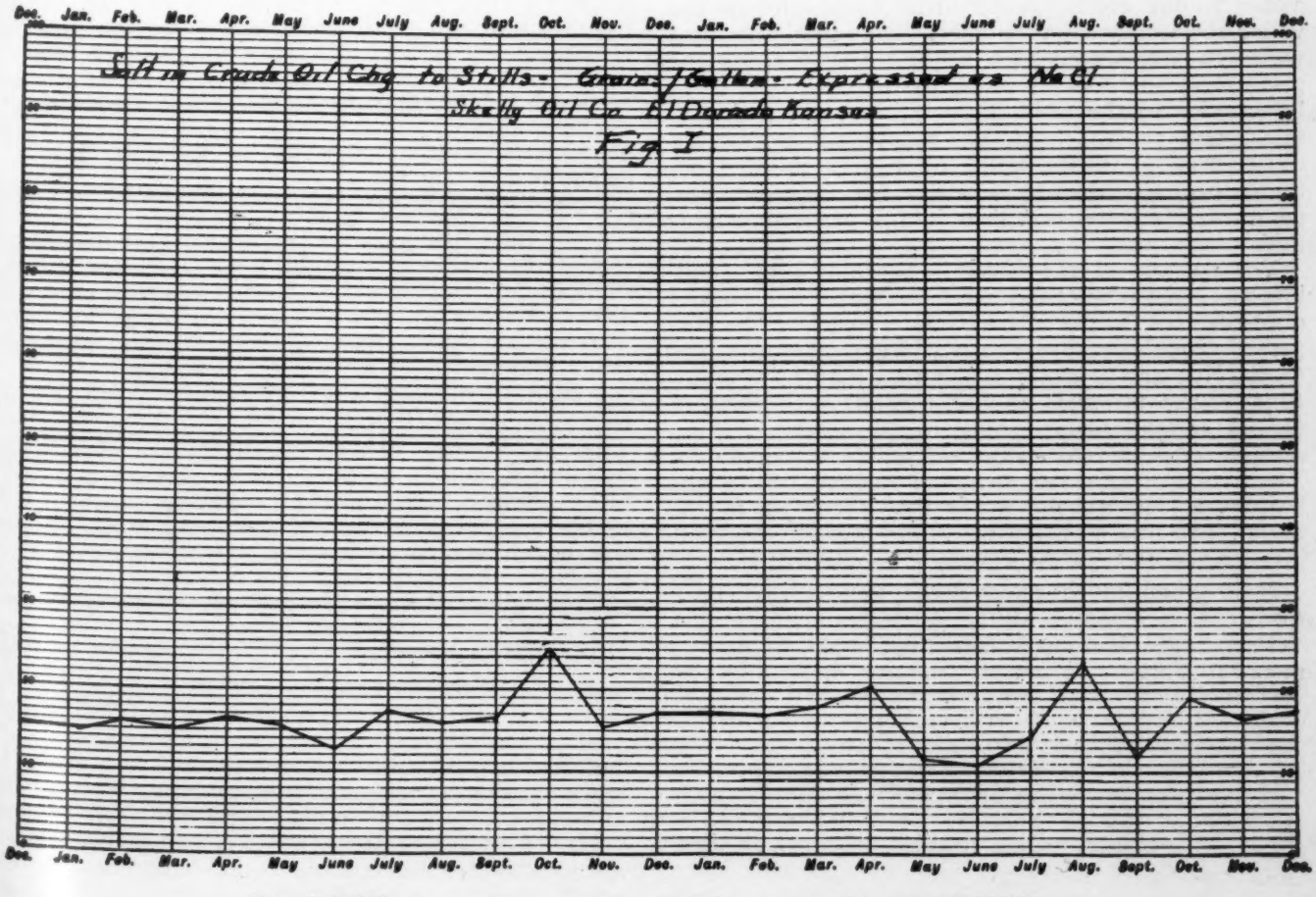
Dec. Jan. Feb. Mar. Apr. May June July Aug. Sept. Oct. Nov. Dec. Jan. Feb. Mar. Apr. May June July Aug. Sept. Oct. Nov. Dec.
Salt in Crude Oil Chg to Stills - Grains/Gallon - Expressed as Na Cl
Skelly Oil Co. El Dorado Kansas
Fig. I
Year of 1935
Year of 1936

FIGURE 1

tact each other and form a globule large enough to separate from the oil.

This oil film can be broken by several methods, but one of the most popular ways is the use of heat. This method is used by The Pure Oil Company in the Michigan Fields[3]. The emulsion is heated to approximately 100°F. and allowed to separate. After the emulsion has become more stabilized, it is often necessary to use higher temperatures. In some instances it is better to heat the mixture to 190-200°F.

The use of temperature has at least three beneficial effects. It lowers the surface tension of the oil, allowing the water particles to coalesce, it lowers the viscosity of the oil, and also causes a greater difference in specific gravity between the oil and brine, thus allowing better and faster separation.

Since this method does not bring about complete dehydration, and since some salt is present in the crystalline form it is necessary to add fresh water to the incoming crude. If sufficient heat is then put on the mixture and the two components intimately mixed, the fresh water will dilute the salt solution already in the oil and will dissolve any salt crystals that are present. The small amount of water that is not removed will be only slightly brackish. The amount of salt dissolved in the water left in the crude can be controlled by the volume of water used and this is one factor that will control the percent of salt removal that is obtained within certain limits.

Some foreign substances will tend to break the restraining film around the water particles, and cause faster separation. Soda ash is one. The raw crude often contains a small amount of free acid which can be neutralized with the soda ash. The soda ash is introduced as a solution in the fresh water that is added to the crude. Water from the outlet of the vapor condensers may be used. This is not the ideal water to use, water such as boiler feed would be preferable, but it is practical, particularly if it flows by gravity to the soda-ash solution tank, and contains some process heat. The solution is introduced in the suction of the crude oil charge pump from there through exchangers where the required temperature is reacted and then into a vapor-separating tank.

The purpose of the tower is twofold: First to allow more time for contact of fresh water and oil and give the fresh water more chance to bring the dispersed globules of brine into a larger mass and also to dissolve any crystalline salt that may be present. Finally, the tower is to bring about separation of vapors and oil before going to the settler. A constant level is automatically held within certain limits. There is also a water draw-off connection on the bottom to remove the water that separates in the tower. The time in the tower under ordinary operating conditions must be carefully controlled.

It is essential to remove the vapors that have formed before going to the settler, for the vapors have served their usefulness and now become a menace. Vapors, if present in any more quantity than a trace, tend to keep the contents of the settler agitated. This impairs separation of water and oil.

The water-oil mixture enters the horizontal settler near one end. The incoming pipe extends up into the settler about one-fourth way and is turned with the outlet opening facing that end of the tank. In this manner the incoming stream is not permitted to flow at high velocity across the main body of oil and as a result the tendency to form eddy currents is reduced. The oil flows out the top at the opposite end, and the water is drawn off the bottom.

The following are some operating data:

	No. 1	No. 2	No. 3	No. 4	No. 5
Salt in raw crude, grs./gal.	19.0	14.1	20.1	20.1	...
BS and W in raw crude, percent	0.6	0.4	0.4	0.3	0.3
Water in mixture, percent	1.7	1.6	1.8	1.5	1.8
Salt in crude after settler, grs./gal.	6.0	4.2	8.1	11.1	...
Water in crude after settler, percent.	0.5	0.3	0.4	0.7	0.3
Salt removal, percent	68.5	70.2	59.7	45.0	...

Data No. 1, 2 and 3 are taken under normal operating conditions, No. 4 is to show the effect of lower temperatures. In this case the temperature on the crude-water mixture was 145°F. The dehydration and desalting were not as good as at higher temperatures. The No. 5 set of data were taken when conditions were such that the time in the settler was nearly 50 minutes and indicates that complete dehydration cannot be expected. This can be explained in this manner. As a majority of the small droplets coalesce, the remaining droplets are, as a result, fewer in number and farther apart, and consequently their chances of contacting and coalescing become less and continue to do so as the process proceeds.

The crude exchangers have given very little trouble since this method has been in use. One refiner[4] that experienced considerable trouble with exchangers salting up has found that if the crude is completely dehydrated before entering the exchangers the salt will not deposit in the tubes.

Since removing a portion of the salt, the necessary amount of ammonia to hold down corrosion in the condensing equipment has been noticeably reduced. The crude heaters make better runs when the salt is removed. As to corrosion of tubes in cracking stills, that will have to be left for further study.

I wish to thank Mr. H. L. Bedell of the White Eagle Division of the Socony-Vacuum Company at Augusta, Kansas, Mr. C. W. Berger of the Globe Refining Company at McPherson, Kansas, and Mr. C. O. Orrison of the Dickey Refining Company at McPherson, Kansas for the information they gave and which was used herein.

[1] Refiner and Natural Gasoline Manufacturer—March 1937, p. 99. The Oil Corrosion Problem—Gustaf Egloff.
[2] Dickey Refining Company, McPherson, Kansas.
[3] The Petroleum Engineer—March 1937, p. 47. Separation and Disposal of Salt Water in Michigan Fields—J. C. Albright.
[4] Globe Refining Company—McPherson, Kansas.

The Extraction of Salts

From Refinery Charging Stocks By the

Electrical Process

G. B. HANSON
Petroleum Rectifying Company

ALTHOUGH salts in charging stocks have long been recognized as the source of many of the troubles encountered in refineries, it is only recently that the problem has become one of major importance. This is due in large measure to the increased use of cracking. Whereas the effects of salts were not particularly serious in skimming and topping operations, they have become more aggravating as the temperatures and pressures used in cracking processes have been increased.

Refiners have been very active during the past few years in efforts to find a practical, efficient and economical method of dealing with this problem. Of the various methods now being employed, the most important are the removal of salts from charging stocks by various means, the use of neutralizing agents to combat the corrosive effects of salts and the utilization of new corrosion-resisting alloys in distillation and cracking equipment.

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W P R A

THE most effective means thus far developed is the extraction of salts from the crude oil before the oil enters the refinery system. This method attacks the problem at its source and has the advantage of removing the salts before they have an opportunity to affect operations or do any damage to distillation equipment. It is in this manner that the electrical process is used.

The principal effects of salty charging stocks upon refinery operation may be summarized briefly as follows: (a) corrosion, (b) coke deposition in cracking equipment, (c) salting and scaling in fractionating and condensing equipment, and (d) the deleterious effect of the salts upon refined and residual products.

The salts responsible for these difficulties are in the crude oils delivered to the refinery and will be found either as crystals or in solution in the residual oil field brine. Large volumes of salts may be present in solution in very small percentages of dispersed water. As an example, for every 3000 barrels of charging stock containing as much as .5 percent of average oil-field brine of about 1.10 specific gravity, 1000 pounds of salt will be taken into the refinery system, all this salt being contained in the relatively small amount of water.

The treatment of limestone formations with hydrochloric acid to increase production has added to the salt problem in some localities due to the formation of salts which become suspended in the crude oil.

The salts that cause most of the trouble in refineries are calcium and magnesium chloride. These salts are hydrolyzed during the distillation process and hydrochloric acid is formed. It has been estimated that corrosion difficulties, due to this cause, will be encountered in the cracking operation if the magnesium and calcium chloride content is in excess of 25 pounds per 1000 barrels (11.4 grams per barrel) of charging stock.

CORROSIVE CYCLE

Corrosion will be particularly severe if hydrogen sulphide as well as hydrochloric acid is present. This is probably due to the fact that the hydrogen sulphide reacts with the metal equipment to form iron sulphide, which in turn reacts with the hydrochloric acid to form iron chloride, at the same time regenerating hydrogen sulphide. This corrosive cycle can be prevented by the removal of the salts in the first instance.

With the increased demand for petroleum products, on-stream time is becoming a factor of considerable importance in refinery operations and existing facilities are being taxed to the limit to meet requirements.

In many instances the length of time a refinery unit can remain in continuous operation is fixed by the rate of coke deposition in the cracking equipment. That this deposition is due largely to the presence of salts in the charging stock is proven by the fact that a reduction in salt content will be directly reflected in the amount of coke deposited. As a result of almost complete salt removal by the electric process, the on-stream time of some cracking units has been increased as much as 1000 percent.

The necessity of shutting down distillation and heat exchange equipment for cleaning has also been reduced as a result of efficient salt extraction. In

addition to the cost of labor involved in cleaning operations of this kind, the resultant wear and tear on equipment is a factor of considerable importance in refinery operating expense.

The quality of residual products will also reflect the advantages of salt removal. This is particularly true of asphalt in which case the presence of salts and other insoluble materials is especially objectionable.

One of the most essential characteristics of a satisfactory refinery desalting process is that it can be incorporated into the refinery system in such a manner that the charging stock may be treated and returned to the system without prolonged settling. The electrical process is particularly advantageous in this respect due to the fact that it involves no separate settling step, the separation taking place within the electric unit. Desalted oil leaves the top of the electric treater while the water containing the extracted salts is continuously withdrawn from the lower portion. The oil passes directly from the treater into a surge drum from which it continues through the refinery system.

The successful operation of the first commercial electrical desalting plant has resulted in the adoption of the process at numerous refineries in various parts of the United States. From the operation of these units, a considerable amount of information has been obtained regarding the efficiency of the process. The results of these operations are briefly summarized in this paper.

OPERATION OF ELECTRIC TREATER

The electrical desalting process includes the use of an electric treater together with an arrangement for obtaining a suitable mixture of the oil with a quantity of fresh water. The desired mixing action usually is obtained by pumping the oil and water through a suitable mixing valve. The pressure maintained on this valve will depend upon the type of oil being processed and various other factors. After leaving the valve the mixture is immediately run into an electric treater where it is subjected to a high-potential electric field, the effect of which is of particular importance in the process. As a result of the action of this field it is found that when the water subsequently settles out of the oil by gravity it carries practically all of the salts originally present. Comparative tests show that a much higher percentage of salt is extracted in this manner than when the mixture is separated by other means. The entire process is continuous and very rapid. The effective salt removal is about 90 percent in every case and in some plants as high as 98 percent. The quantity of wash water required varies from 5 to 20 percent of the volume of oil.

The operating temperature varies from 130 to 200° F., depending upon the nature of the oil and local conditions. In all cases the temperature must not be high enough to start vaporization inside the treater tank. The treater tanks are 10 feet in diameter by 12 feet high and designed to operate at 35 pounds per square inch. If desired, back pressure can be maintained on the tanks by use of valves installed on the outlets. The power consumption will run from 1.5 to 2.5 KW. per unit. The maximum demand per unit will be 8 KVA but this demand will seldom occur.

The electrodes are so designed that the conducting surfaces are very small and the leakage of power through the oil is kept at a minimum.

Transformers rated at 5 KVA. giving potentials of about 16,500 volts are commonly used in desalting work, and when two transformers are employed on a single desalting unit, as is usually the case, the circuit is designed so that the high-tension side of each transformer is connected to one of the electrodes.

The polarity of the transformers is so arranged that when connected in this manner there will be a potential of 33,000 volts between the electrodes but only 16,500 volts from each electrode to the grounded treater shell. This results in localizing the maximum voltage in the treating field where it is needed and simplifies the problem of insulation.

An automatic switch at the top of the treater serves to control the power to the unit. When the liquid level is lowered for any reason the power is cut off. Electricity thus can be delivered to the electrode system only when the treater is full of liquid.

The results obtained in operating the various installations are remarkably uniform regardless of the type of oil being handled. Capacities per unit range between 1500 and 2000 barrels of oil per day, and the reduction of salt has averaged over 90 percent.

Some of the effects on refinery operation thus far noted are:

(1) Improvement in asphalt quality.
(2) Longer and more uniform operating periods.
(3) Elimination of corrosion.
(4) Saving in fuel.

Citation of results obtained in a few commercial installations may be of interest in showing capabilities of the electric desalting process.

In the case of one installation for desalting 36 gravity A.P.I. Eastern Kentucky oil the crude contains between .5 percent and 1 percent B.S. with a salt content varying between 100 and 300 grams per barrel, consisting principally of sodium, calcium and magnesium chlorides. The charging stock consists in part of fresh production and partly of oil that had been in storage for several years. The salts present in this oil are of such character and are present in such quantities that before installation of the electrical desalting process the asphalt manufactured at this refinery did not meet solubility specifications. Installation of the new desalting equipment increased the solubility of the asphalt from 99.50 to 99.95 percent, or better, and reduced ammonia requirements to a negligible amount. Desalting also practically eliminated previously experienced corrosion of equipment and increased the length of possible runs of the cracking units by over 300 percent.

When the refining equipment is shut down for cleaning it has been found that the coke is soft and easily removable as distinguished from being very hard and difficult to remove before installation of the electrical desalting equipment.

This plant is now being operated at a temperature of about 180°F. and reduces the salt content from an average of 168 grams per barrel to 10 grams per barrel.

PROLONGED OPERATING PERIOD

In another refinery producing large quantities of asphalt it was found that while solubility requirements could be met, the oil contained so much salt that economical operation of the asphalt unit could not be maintained, and it was necessary to shut down for cleaning this unit every few days. The plant was also experiencing considerable corrosion trouble and operated on oil of approximately 18° A.P.I. gravity containing salts up to 150 grams per barrel. After installation of an electric desalting unit these troubles

were largely eliminated and the plant has been in operation for over one year with a daily capacity between 2500 and 3000 barrels. Due to the resulting efficient operation and long uniform operating periods, this refinery unit is now turning out an exceptionally high grade of asphalt. At the same time, the asphalt plant can be kept in operation for 60 days or more without the necessity of shutting down.

This same refinery has a second electric-desalting plant in operation on another refining unit treating 24° A.P.I. gravity oil containing as high as 80 grams of salt per barrel.

Both these plants are operating at approximately 160° to 175° F. and are reducing the salt content to less than 8 grams per barrel. The reduction in hydrochloric acid generation in the cracking units is approximately 99 percent.

Another plant was installed for the purpose of handling a 40-gravity Michigan crude. This oil contained up to 120 grams of salt per barrels, and, since an extraordinarily large percentage of this salt was magnesium chloride, the corrosive effects of this oil were unusually severe. This refinery does not make gasoline but turns out a number of specialty products and sells the reduced crude to another refiner for cracking purposes. The salt content of the original crude is being reduced to about 2 grams per barrel. The generation of hydrochloric acid in the topping unit has been almost completely eliminated, and the salt content of the reduced crude has been lowered to such a point that the cracking unit is able to run many times longer than before installation of the electric desalting equipment. In addition to the salts found in the oil, the equipment at this plant also removes a large quantity of solid matter which is in suspension. The run-time of this unit was increased from an average of 12 days to an indefinite period. One continuous run of over 100 days was reported.

The same company installed a 6000-barrel electrical desalting plant on a combination high-pressure unit which had just been built at another of its refineries. The run-time of this unit was increased from 7 to over 30 days by installation of the desalting equipment. During the operation of this plant a 7-day test run was made on a blend of oils from the Mid-Continent fields, principally Seminole production. This test was run for the purpose of obtaining data for the design of a second refinery unit to handle this crude, and the results obtained with the desalting equipment were so satisfactory that this equipment has been incorporated in the design of three new refining units which this company is building.

A two-unit installation was made for a 2500-barrel refinery in Kansas. This refiner was desirous of carrying out a viscosity-breaking operation but on account of the salt present in the oil, the furnace plugged up after about 7 hours operation. The electric desalting units reduced the salt to less than 4 grams per barrel and enabled the refinery to run its viscosity breaker for an indefinite period. Furthermore, all corrosion was eliminated. As a result of experiments it was found possible to reduce the water requirements to 125 barrels per day for 2500 barrels of oil.

WIDE VARIETY OF OILS

These installations, operating on such a wide variety of oils ranging from 18 to 40 gravity and including four general classes of oils being produced in the Mid-Continent, indicate that the process is applicable to nearly every type of crude being processed. Samples of oil from many points throughout the Mid-Continent area have been tested and many of these oils show extremely high salt contents, although they have been treated in the field to a low-water content. Samples obtained from East Texas fields show that the salt content of this oil is rapidly increasing and that, as the water table rises in this field, the refiners handling East Texas crude may expect more and more trouble from salt. Rodessa crude has been found to contain approximately 35 grams of salt for each one-tenth of 1 percent B.S.&W. content. Samples from the coastal fields of Louisiana have been found to contain as high as 600 grams per barrels. Many other samples from all parts of the Mid-Continent and Gulf Coast fields, as well as from abroad, indicate that, regardless of the source of the oil, there is nearly always an appreciable amount of salt present. Most of the samples were obtained from oils which have undergone a reasonable amount of settling, thus indicating that the percentage of W. & B. S. had been reduced to the lowest value which could be expected.

Analyses of the brine from these oils indicate the presence in nearly all cases of enough magnesium chloride to give an appreciable evolution of hydrochloric acid on heating. Samples from Urbana, Arkansas; Michigan; Duncan, Seminole, Cushing, and Drumright, Oklahoma: Port Barre, Louisiana; White Castle, Louisiana; and Porterville, Texas, indicate that plants handling these oils will not only have corrosion problems but can expect a certain amount of stoppage of the equipment, due to the settling out of the salt. It is also becoming noticeable that the quantity of salt is increasing throughout the Mid-Continent area, and this fact, together with the increasing number of modern refineries being installed, makes the problem assume very serious aspects. One refinery has found that a new 20,000-barrel unit recently completed will handle only certain portions of the oil available and that it is necessary to keep old shell-still equipment in operation until the problem of salt removal can be solved. Another refinery handling 10,000 barrels per day of Rodessa crude gets between 50 and 200 hours operation on the topping unit and about 24 hours on the subsequent cracking operation. Naphtha from the topping unit is highly corrosive and solid matter settling out in the furnace tubes averages 42 percent salt.

Operating costs include power for the treaters and pumps, the cost of water, some heating expense depending on equipment, and the process royalty. On plants installed to date no extra labor has been required. The operators inspect and regulate the treaters along with their other duties. Most refiners have not charged heat against the cost of electrically desalting, due to the fact that the only additional heat required has been that necessary to raise the temperature of the wash water. In such installations where the condenser water is reasonably soft, it has been the practice to take hot water off the condenser boxes, and this water requires very little additional temperature for satisfactory desalting. Power costs will average about one mill per barrel.

The chief advantages of the electric method are its efficiency, simplicity and reliability. The removal of the salts will be 90 percent or better on nearly every type of oil. The operation is semi-automatic and gives ample warning of impending process failure so that operation can be corrected before the refinery proper is affected. The process will handle every type of reasonably fresh crude encountered to date, and has occasionally handled storage several years old.

The Stratco PbS Process

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THE authors discuss the cost of several treating processes as well as advantages of a number of licensed systems. The paper was presented before the technical session of the twenty-fifth annual meeting of the Western Petroleum Refiners Association, Hot Springs, Arkansas, April 26-28, 1937. The statements and opinions expressed therein are those of the authors.

W P R A

THE sulfur present in raw gasolines may occur in the free or elementary state, some as hydrogen sulfide, but the largest amount is present in the form of organic derivatives. This latter group is divisible into two classes, (1) the acidic, and (2) the neutral. The first class includes mercaptans, which, together with hydrogen sulfide, are mainly responsible for the unpleasant odor, and positive reaction to the "doctor" test. The second class includes the organic disulfides, sulfides, and thiophenes, which in general are sweet to the "doctor" test.

Hydrogen sulfide, if present in the gasoline, is removed easily by applying caustic soda solution, since it is a comparatively strong acid and either or both of the hydrogen atoms may be replaced to form metallic sulfides which are water soluble. However, mercaptans are relatively weak acids and sodium mercaptides are hydrolized readily as shown by the reversible reaction: $RSH + NaOH \rightleftarrows RSNa + H_2O$. In other words, unlike hydrogen sulfide, the removal of mercaptans from gasoline by caustic soda solutions is extremely difficult to complete, and other sweetening methods had to be found.

Refiners at an early date turned to the now well-known sodium plumbite and sulfur (doctor) method of treatment in which mercaptans are converted to disulfides through the media of lead mercaptide formation, sulphur being used to oxidize the lead mercaptides to the corresponding organic disulfide and lead sulfide. In order to reduce chemical costs lead sulfide is then regenerated to sodium plumbite. It was soon noticed in commercial practice that the use of regenerated doctor solution frequently resulted in increased ease of treatment, and further, that the explanation of this effect was the presence of some non-converted lead sulfide in the regenerated doctor solution. This discovery resulted in the development of several methods of lead sulfide treatment. In its earlier form the method was to contact the sour gasoline with a suspension of lead sulfide in sodium plumbite solution, completing the reaction by adding sulfur as usual. Later, Rowsey[1] found sweetening could be accomplished by contacting the sour gasoline with a suspension of lead sulfide in caustic soda solution while blowing air through the mixture; in many cases it was necessary to add sulfur to complete the reaction. The inherent disadvantages of these earlier methods are discussed below.

METHOD OF OPERATION

In attempting to sweeten many gasolines, particularly cracked, using a suspension of lead sulfide in caustic soda, employing no sulfur, it was found that the gasoline suffered a severe darkening and the reagent developed a tendency to carry-over with the sweetened gasoline resulting in large chemical losses. It was determined that these difficulties were attributable to the formation of undesirable concentrations of soluble lead salts within the reagent by the action of the process air on the lead sulfide. Reprecipitation of the lead compounds as lead sulfide was found to eliminate this difficulty.

In the improved PbS method of sweetening, therefore, sour gasoline is intimately contacted with preformed dense and quick-setting lead sulfide suspended in strong caustic soda solution. To this mixture is added a small positively regulated quantity of air for the purpose of activating the reagent; also a sufficient quantity of aqueous sodium sulfide solution is fed continuously to the mixture for the purpose of precipitating all excess soluble lead salts that are formed in the sweetening operation. Thus, the lead sulfide reagent is maintained in a desirable chemical balance, and the sweetening operation is accomplished essentially as a catalytic oxidation under controlled conditions. In this manner consumption of treating chemicals by side reactions is reduced to a minimum.

The operating procedure may be seen readily by reference to the typical flow diagram, Figure 1. Lead sulfide reagent and sour gasoline, together with small volumes of air are fed into the system through the contacting device. The amounts of the latter depend upon the sourness of the raw gasoline. These intimately mixed constituents pass to a horizontal settler where the reagent settles out. The reagent recirculation pump takes suction on the reagent settled out in the bottom of the settler, and discharges to the contactor; the volume of reagent recirculated varies from 10 percent to 50 percent, or higher, again depending upon the sourness of the gasoline. The sweet gasoline, containing a very small amount of caustic and reagent in suspension, then passes through the rock-packed clarifier and thence to storage. Flow controllers on the feed, air and sodium sulfide make for ease and utmost simplicity of operation and greatest economy of chemicals.

The most important feature of the operation (and that upon which the improved process is based) is the maintenance of the reagent in a substantially balanced condition by continuous addition of the correct volume of sodium sulfide solution. This is most readily accomplished by the flow controller.

MAINTAINING STRENGTH

Second only in importance is maintaining the strength of the caustic soda solution. This solution loses strength

gradually owing to the addition of the aqueous sodium sulfide solution and the formation of water by the sweetening reactions. It has been found that when the caustic strength falls below a certain minimum, depending upon the character of the stock being treated, treatment becomes more difficult and the tendency of the PbS particles to pass out of the settler with the sweet gasoline increases, resulting ultimately, if allowed to continue, in carry-over of reagent to finished gasoline storage. At such times the caustic portion of the reagent is removed from the system and replaced with stronger caustic solution and the sweetening operation is resumed. The operating cycle in this respect varies from 10 days to 3 months, depending upon the character of the gasoline being sweetened and the volume of sodium sulfide solution added. The weak soda solution may be used for hydrogen sulfide removal or elsewhere in the refinery where caustic is required.

It has been found also that after long continued use the reagent decreases in activity, probably because the PbS particles in the reagent become coated with a film. The reagent is then removed from the system and revivified. This procedure consists merely in washing the lead sulfide with hot water and steam for a short time which results in removal of foreign bodies constituting the film. The clean lead sulfide particles drop to the bottom of the washing vessel, are mixed with strong caustic, and returned to the system for use. The frequency with which the PbS requires washing depends upon the characteristics of the sour gasoline—generally from 3 to 12 weeks.

Process air requirements are very low, in general from nil to about 0.15 cubic feet per barrel except with high mercaptan content stocks where air requirements may reach a maximum of 1.0 cubic feet per barrel. The nitrogen and any excess air are removed from the system through an inverted steam trap installed on the settler. This gas may be returned to a vapor-recovery system although experimental work has indicated that the amount of hydrocarbon carried by the gas is almost negligible.

The finely divided PbS and caustic, carried in suspension, in the sweet gasoline entering the clarifier are removed by the coalescing action of the rock, and flow to the bottom of the vessel. The small amount of reagent, accumulating in the bottom of this vessel, is removed about once every 8 to 16 hours and returned to the system.

The several PbS sweetening plants now in operation either run continuously or do the required amount of treating in a part of an 8-hour shift and shut down for the balance of the day. Substantially no adverse effect has been noted during discontinuous operation, as compared to continuous.

The flow diagram does not show the three small tanks normally used for storage of sodium plumbite and sodium sulfide solutions, and reagent preparation (or washing) respectively. Fresh PbS reagent is made by adding the required amount of sodium sulfide solution to sodium plumbite solution so that all of the lead is precipitated out of the latter.

2. OPERATING RESULTS

Large scale commercial operation of the improved lead sulfide-sweetening process over a period of 6 years has demonstrated the general practicability of the method for treating both straight-run and cracked gasoline and also heavier products. These operations have provided ample opportunity to obtain reliable data on yields, chemical consumption, and effects on quality of the finished products.

Following are results secured during normal commercial operation using the PbS process for sweetening sour cracked gasoline:

1. *Chemical consumption*—pounds per barrel of gasoline.*

The following figures represent maximum requirements:

Percent mercaptan sulfur†	0.01	0.02	0.03
Sodium Sulfide	0.03	0.06	0.09
Litharge	0.0077	0.0085	0.01
Caustic Soda‡	0.0077	0.0085	0.01

2. *General*

(1) Low octane losses.

(2) color drop—nil to three points depending upon crude source endpoint of gasoline, etc.

(3) Satisfactory color stability.

(4) Volumetric loss—not measurable with the exception of that due to evaporation in tankage.

(5) Sulfur—no increase during sweetening.

(6) Copper corrosion test—unchanged or improved.

3. *Use of Air for Sweetening.* Considerable interest has been displayed relative to the effect of process air additions during sweetening on the gum content and gum stability of the finished gasoline. It has been found that, within the limits of error of the determination,

* For cracked gasoline and straight-run gasoline containing a lower percentage of mercaptan sulfur than indicated chemical consumption is always less.

† Borgstrom & Reid Method.

‡ Net consumption of caustic after credit is given for diluted caustic withdrawn from the system.

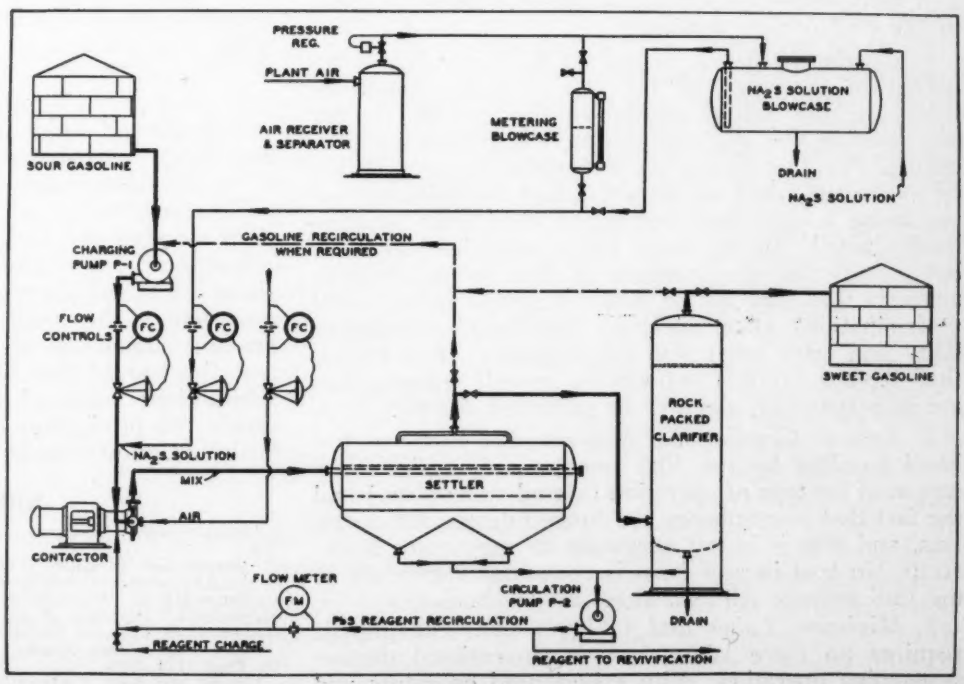

FIGURE 1
Typical Flow Sheet—Stratco PbS Sweetening System

the amount of air added is so small that its effect on the gum content and stability is negligible.

3. ADVANTAGES OF STRATCO PbS PROCESS

From a study of the available data on sweetening stocks of similar quality in plants where the PbS process replaced the conventional doctor method with efficient revivification procedure, it was indicated that the advantages of the PbS process were approximately as follows:

1. Decreased chemical costs—of the order of 50 percent or more.
2. Decreased octane loss—of the order of 0.5 points CFRMM.
3. Decreased color loss.
4. Decreased volumetric loss—as high as 1.5 percent.
5. Decreased utilities cost—principally due to eliminating regeneration of blackstrap, and avoiding water washing of sweetened gasolines.

Unit costs for chemicals will vary considerably between individual plants, as will figures for stock value, octane value, and utilities but it will be apparent that the potential savings to be realized represent a considerable sum for all plants.

A considerable number of cracked and straight-run gasolines, commercially produced from a wide variety of crude sources, has been examined recently. For purposes of comparison these stocks have been sweetened by the conventional doctor and PbS processes, and in a number of cases, by other less-known sweetening methods. A study of the results and review of the processes involved indicated that many of the following advantages can be realized by PbS sweetening as compared to other methods.

1. Minimum chemical costs.
2. Minimum octane degradation.
3. Best color retention.
4. Eliminates the use of sulfur.
5. Negligible losses of gasoline.
6. No corrosive chemicals used.
7. Minimum labor and utilities cost.
8. Continuous and automatic operation.
9. Unusually low plant cost.
10. Inexpensive conversion of doctor plants.

A more detailed discussion of some of the advantages of the PbS process follows:

1. *Eliminates the use of sulfur.* No free sulfur is required in carrying out the PbS treating operation; therefore, the sulfur content of the gasoline is not increased in the slightest during the sweetening procedure. For this reason the difficulties with regulation of the sulfur freed are removed, and the possibility of producing a corrosive (bad copper-strip test) gasoline is eliminated. Furthermore, recent experimental work has shown that the presence of free sulfur, even in amounts that may not be detected readily in the usual determinations, affect adversely both inhibitor susceptibility and tetra ethyl lead requirements. It is known that reports on this subject by several investigators are in preparation and will be published shortly.

2. *Losses—Gasoline and Reagent.* The losses of finished gasoline by the PbS process are negligible by reason of the type of operation (closed system, etc.) and the fact that no emulsions are formed during the operation, and that it is not necessary to regenerate blackstrap. No lead in any form is consumed chemically in the PbS process. All lead losses are mechanical.

3. *Minimum Labor and Utility Costs.* The process requires no more labor than a conventional doctor-sweetening operation. With experienced operators and with automatic controls, the labor and supervision required usually are materially less than in a doctor plant.

The cost of all utilities required by the PbS process is decidedly less than for conventional doctor sweetening. No air is required for external revivification, and very little steam. In contra-distinction the cost of steam and air required for blackstrap regeneration has often been found to be nearly equivalent to the cost of chemicals required in the sweetening operation. Furthermore, in substantially all sweetening processes, the sweetened gasoline is washed with water or chemical solutions prior to leaving the plant. This final wash is eliminated in the PbS process.

4. *Continuous and Automatic Operation.* The operation of the PbS process is continuous and automatic. Consequently the consumption of chemicals can be readily held to a minimum, and chemical control is reduced to the utmost simplicity.

5. *Plant Costs — Conversion of Existing Doctor Plants.* It will be noted from the flow sheet that the design of plants utilizing the PbS process is extremely simple and that existing doctor plants can be converted to the process cheaply and without difficulty. In practically all cases, most of the equipment can be utilized in making the conversion.

4. MECHANISM OF REACTIONS

In conclusion, the question of mechanism of the PbS sweetening reaction has arisen many times. It appears that there are three schools of thought on the subject. Briefly, they are:

1. *Variations of the well-known doctor treating reaction.* The basic reactions of the conventional doctor treating process and blackstrap revivification are well known. Drs. Morrell and Faragher[2] have published several articles of outstanding merit in this connection, presenting data to indicate that PbS treating is a special case of doctor treating. Ott and Reid[3] have a different hypothesis, considering sodium plumbite as lead hydroxide and show experimental data indicating that both basic and neutral lead mercaptides may be formed. These may be oxidized either with sodium sulfide or free sulfur to form the disulfides.

2. *Catalytic Oxidation.* Wendt and Diggs[4] in an early article assigned a catalytic role to the suspension of PbS in caustic, the lead sulfide not entering the reaction but acting as a catalyst to promote the oxidation reaction $4RSH + O_2 \rightarrow 2R_2S_2 + 2H_2O$.

3. *Adsorption.* Wendt and Diggs in a later article, and Slagle and Reid[5] in a recent publication, reported that PbS was capable of adsorbing more than 25 percent by weight of mercaptans.

However, though all of the various hypotheses explain some of the findings noted during several years of commercial operation, none of them individually explains all of the phenomena. It is believed that an explanation of the mechanism of the reaction probably lies in a summation of all three. The important fact is that the use of the Stratco PbS process, employing substantially balanced reagant, on sour gasoline will insure the production of a sweet gasoline of highest quality at an extremely low overall cost.

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The Economics of Treating and Inhibiting Gasoline

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IN manufacturing motor fuel for internal combustion engines, the refiner must keep in mind two paramount factors. The first is to deliver a final product of suitable quality and the second is to do this at lowest possible cost. While it may be reversing the proper order of things, the importance of the preservation of quality justifies the mention of it before giving the details of just what constitutes quality. It is rather obvious that once a gasoline has been made, there should be as little change as possible in its quality from the time it leaves the refinery until final consumption. The criteria of quality of varying importance are gum content (potential or dissolved), stability of gum characteristics as determined by resistance to oxygen under pressure, sweetness or odor, freedom from corrosive characteristics, sulfur content, vapor pressure, octane number, color and distillation range. Ordinarily when speaking of treating gasoline, certain of the characteristics as listed above are excluded. This applies particularly to vapor pressure and distillation range which are, or should be with proper facilities, little affected by treating. For this reason our discussion will not deal with this part of gasoline quality.

The writers of this paper do not intend to set up what constitutes proper criteria of quality as regards any of the above items. The individual refiner will have to determine these for himself, based on many factors with which he alone is entirely familiar. It naturally follows that each refiner must really determine for himself, based on what constitutes adequate quality for his materials, just how these properties may be most economically attained. Such being the case, it is hoped that the differences between the experience of others and that described in this paper will be interpreted as normal variation in a rather complicated manufacturing process.

A further difficulty which many times in the past has obscured the proper valuation of refining processes, is that methods of cost accounting have not been standardized in the petroleum industry. One part of this accounting procedure which has been at times almost entirely neglected but which is more recently receiving proper attention, is the matter of degradation of one quality in processing the gasoline to attain another. This applies principally to the question of octane values. For many years the average refiner has had no idea just what doctor sweetening was costing him because he did not appreciate the effect of treating with doctor solution and sulfur upon octane value and more recently on tetraethyl lead susceptibility.

The discussion below may be of some assistance to the individual refiner in checking over his present methods of processing, and perhaps with the aid of some experimental work on his own products, to arrive at the most economic handling of gasoline in his refinery. If this be true, the purpose of the paper will have been achieved even though the answer to a particular set of conditions is not contained in it.

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W P R A

TWO CLASSIFICATIONS

The treating of gasoline may be roughly sub-divided into two classifications, though very often a single refining process will take care of both. To achieve proper color, gum content, or stability of certain properties in storage, the raw product in question may be treated with sulfuric acid, an absorbent, or by miscellaneous materials. Included in this category should be addition of an anti-oxidant or inhibitor which may be used by itself to deliver a satisfactory product, or combined with any of the processes mentioned above.

The second classification has, by a combination of circumstances, achieved an importance in the treating of gasoline which is not entirely consistent with its real value. Reference is made to the sweetening of gasoline, primarily a treatment to impart satisfactory odor to the finished product.

In treating gasoline under the first classification, the simplest and most economical method is the addition of anti-oxidant to either sweetened or washed gasoline. Whether or not a merchantable gasoline can be made by sweetening or washing and inhibiting depends, of course, on the nature of the material being handled and the quality desired. The equipment necessary for the addition of the inhibitor is not expensive for the reason that the inhibitor may be injected into the gasoline in the blending operation. The equipment to handle 10 or 15 thousand barrels of gasoline per 24 hours can be installed at a cost of $2,000 to $3,000 at the most. Some figures available to the authors and given in Table I indicate a total cost of $0.0026 to $0.0227 per barrel treated for the entire operation, including the anti-oxidant. The cost of the anti-oxidant is from 70 to 90 percent of the total cost. Table I shows the details of the gum and induction period tests together with the cost of inhibiting of 6 commercial gasolines. The figures give the average for a considerable period of operation. In all cases the low-pressure cracked gasoline has been treated with an absorbent.

If all gasoline could be finished thus, there would be no problem to discuss. It must be borne in mind that inhibiting involves no loss of gasoline or degradation in quality. Octane number or lead susceptibility is totally unaffected initially. The only precaution which must be taken in the stor-

age of inhibited gasoline—since anti-oxidants as a rule are slightly soluble in water or are attacked by caustic soda and the like—is to avoid loss of inhibitor through contact with these materials. If the proper precautions are taken depending on the initial tests and the amount of inhibitor added to the gasoline, quality, as regards octane and gum tests, will be maintained for an extended period.

TABLE 1

CRUDES	Straight Run	Low Press. Cracked	High Press. Cracked	Casing Head	TESTS BEFORE INHIBITING A.S.T.M.	TESTS BEFORE INHIBITING Copper	TESTS BEFORE INHIBITING I.P.		TESTS AFTER INHIBITING A.S.T.M.	TESTS AFTER INHIBITING Copper	TESTS AFTER INHIBITING I.P.	Cost/ Barrel
1. 41% Mid-Continent...	53.09%	37.95%	.....	8.96%	1.4	35	2.75	Average..	1.25	11.2	6.15	$0.0215
59% Pennsylvania....	.....	.....	.....	.....	0.2–3.0	10–70	1.5–4	Range....	0.2–3.0	7.5–20	6–6.5	.....
2. 6% Pennsylvania.....	36.00%	16.68%	44.54%	2.78%	0.85	50.5	3.25	Average..	0.93	15.7	4.27	0.0026
94% Mid-Continent...	.....	.....	.....	.....	0.8–1.2	36–74	3–3.5	Range....	0.4–2.0	9–19.5	4–4.6	.....
100% Mid-Continent..	48.76%	9.59%	27.80%	13.85%	1.3	53	3.75	Average..	1.11	7.2	4.21	0.0054
	.....	.....	.....	.....	0.6–3.0	12–75	3.5–4.0	Range....	0.2–2.6	2–11.4	4.0–5.0	.....
4. 38.5% Corning.....	46.44%	22.97%	24.84%	5.75%	0.76	64	3.52	Average..	0.63	10.6	5.61	0.0123
61.5% Mid-Continent.	.....	.....	.....	.....	0.4–1.6	21–127	3.0–4.75	Range....	0.4–1.0	8–15	5.2–6.3	.....
5. 56% Mid-Continent...	37.78%	34.91%	27.28%	0.03%	1.92	13.9	3.49	Average..	0.85	9.8	4.78	0.0060
44% Michigan.....	.....	.....	.....	.....	0.5–2.0	9–20	2–4.5	Range....	0.0–2.2	2.6–13.9	4.3–6.0	.....
6. 88.6% Mid-Continent.	48.32%	4.00%	47.65%	0.03%	1.31	34.7	3.0	Average..	1.66	9.4	6.38	0.0227
11.4% Gulf Coast.....	.....	.....	.....	.....	0.4–3.2	12–58.5	2–3.75	Range....	0.8–2.6	4.4–13	5.25–6.75	.....

RESPONSE TO INHIBITORS

However, it has been noted that certain gasolines have either a limited response to inhibitor or practically no response. In addition certain other gasolines have a sufficiently high dissolved-gum content to make a further refining operation necessary for that reason alone. If the desired quality cannot be achieved by the use of anti-oxidant alone, the use of an absorbent material may either give a gasoline, which, followed by sweetening or washing, is a satisfactory product, or one that can be easily made so by the use of an anti-oxidant. The two better known methods, both normally using fullers earth, are the Gray Process and the Osterstrom Process. Both depend on selective polymerization for the refining action. The former operates in the vapor phase; the latter in the liquid phase.

The operation of Gray towers is probably familiar to all refiners and where an end-point gasoline is delivered by a cracking unit, the vapors pass at equilibrium boiling point directly from the final fractionating column through the Gray equipment and after-fractionator, whose purpose is to remove entrained polymer material. This means that the only operating costs involved cover the use of the additional equipment, the renewal of the bed of fullers earth and the royalty. The cost of the equipment (1) will average between $15.00 to $25.00 per barrel daily capacity. The paid up royalty is $35.00 per barrel of daily average production. Some published figures[1] on the operation of a direct connected Gray tower based on the conservative average of 4,000 barrels per ton of fullers earth indicate a cost of $0.0057 per barrel. From the experience of the authors this figure is high and the average cost on low-sulfur stocks principally of Mid-Continent origin, is from $0.002 to $0.004 per barrel. The low operating costs noted are, of course, very important, but more important is the fact that due to the highly selective action of the fullers earth there is no volumetric loss corresponding to that which results by contact with sulfuric acid. A certain amount of the gasoline—from 0.5 to 1.25 percent of material cracked at super atmospheric pressure—is polymerized outside of the gasoline range and eventually finds its way to fuel oil. But according to the claims of The Gray Processes Corporation[1], this loss is possibly offset by secondary polymerization of cracked gases to higher-boiling material falling within the gasoline range. The further claim is made that there is no discernible decrease in octane value of gasoline by this method of handling. Certainly the experience of the authors would substantiate the claim that this loss on high-pressure cracked gasoline is under 1 octane by the A.S.T.M. motor method. With low pressure cracked gasolines of the Gyro type, the operating cost would run higher in proportion to the reduced yield per ton of fullers earth. The loss by polymerization to material outside of the gasoline range may run as high as 3 to 4 percent.

The use of fullers earth under superatmospheric pressure and at elevated temperatures is known to the industry as the Osterstrom process. The limitation in the use of it is occasioned by the fact that it is necessary to condense the make of a cracking unit, pass it through a tube still under considerable pressure, then through a small bed of fullers earth, after which the material is flashed into a fractionating tower. The latter separates the polymers as the bottoms and gives the treated gasoline as the overhead product. The conditions found most advantageous in the industry are a pressure of from 700 to 1000 pounds and a temperature of 475° to 575° F. The cost of installation runs from $22 to $40 per daily barrel of material treated. The royalty is the same as that for the Gray Process, and the operating cost runs from $0.0225 to $0.040 per barrel. The losses, both as to volumetric reduction in gasoline yields and octane, approximate those obtained under the Gray process. The feature which distinguishes the Osterstrom process and makes it advantageous for use with products difficult to treat, is the fact that the extent of polymerization and thus refinement can be controlled by varying the temperature and/or pressure. Usually the variation of the temperature alone is sufficient to give the desired effect.

OTHER REAGENTS

There are numerous other catalytic or chemical reagents which are used to treat gasoline for reduction of gum or attainment of other desired qualities. Reference will be found in the literature to the use of numerous catalysts and reagents, but the only one which has achieved commercial importance is the Lachman process employing a concentrated solution of zinc chloride at elevated temperatures. In this process, a 70 to 85 percent solution of zinc chloride is used at a temperature of from 265° to 410° F., depending on the boiling range of the material undergoing treatment. The solution is

sprayed downward through a packed tower and meets the ascending raw vapors. The refining action is apparently catalytic, causing selective polymerization, although some claim is made that with proper control a product not requiring any further treatment for sweetness can be delivered. It is also stated that reduction in sulfur content can be effected. This is reported to reach 30 to 40 percent[2] of the original sulfur content. The writers do not have any figures available as to installation costs and the only figure as to processing cost is $0.06[3] per barrel on a normal cracked distillate. It is claimed that the solution handling losses in this process is less than 1 percent and the polymerization loss is from 1 to 3 percent, depending on the type of material treated—the higher figure presumably applies to "vapor-phase" cracked stocks. According to published data the octane value is practically unaffected. By this method of treating, gum, as interpreted by the various methods, is reduced and under certain circumstances practically all mercaptan sulfur removed.

The use of sulfuric acid for various refining purposes in the petroleum industry is very thoroughly covered in the literature. It was practically the first chemical reagent used in the industry, certainly to any great extent. Where a high sulfur stock is being handled, and it is found necessary to reduce the content, sulfuric acid is still the only reagent which gives reasonably satisfactory results. A great deal of work has been done in developing this process covering the variables of concentration of acid, time and temperature. However, with more normal cracked stocks it is possible to use processes previously mentioned in this paper and it is on the basis of such stocks that the following discussion is based.

In acid treating low-sulfur gasoline one may use either the traditional batch or continuous treater. The former type of equipment, of course, gives excessive losses and its use in the industry has been practically eliminated for this purpose. For better control the Stratford process[4] has been developed which, by the use of centrifuge equipment, gives closer control on time of contact and separation of sludge. The Halloran process, well known on the West Coast, is particularly applicable to high-sulfur gasolines. For this reason the higher installation and operating costs are not included in our figures.

In all acid treating of gasoline and particularly in the handling of cracked stocks the treated material must be rerun on some type of distillation equipment to give a finished product. In some cases greater economy is obtained by separating by fractionation the lighter ends not requiring treatment. This reduces the quantity of material acid treated and rerun. Equipment for acid treating gasoline costs from $15.00 to $30.00 per barrel daily capacity. Redistillation equipment will range from $35.00 to $50.00 per barrel daily capacity. The treating cost will run from $0.025 to $0.045 per barrel. Depending on the quantity and strength of acid used, the volumetric or solution loss in treating will run from 1.5 to 2.5 percent, and the polymerization loss from 1 to 2 percent. The reduction in octane value of finished gasoline will be on the average of from 2 to 3 octanes by the C. F. R. motor method and the increase in tetraethyl lead requirement is from 0.2 to 0.5 cc. per gallon.

The refining steps described above are chiefly for the purpose of reducing gum, stabilizing the qaulities of the product under treatment, or improving color. As mentioned above, the matter of delivering a so-called sweet product is considered an important part of refining gasoline. The question of sweetening is considered by itself in order to facilitate a direct comparison of costs and losses, even though in acid treating the treatment for reduction of gum and stabilization of properties may be combined with the sweetening process.

Traditionally, gasoline is sweetened by the use of sodium plumbite solution and sulfur. In spite of advances made in the art, probably most of the sweetened gasoline sold in this country is still processed by the use of these reagents. Continuous equipment for doctor sweetening will cost from $10.00 to $20.00 per barrel daily capacity of material handled. Some figures available to the authors indicate a total sweetening cost of from $0.0082 to $0.0216 per barrel. This covers a multiplicity of stocks, mainly of Mid-Continent origin but including some Gulf Coast, Ohio and Michigan crudes. For the straight-run gasolines the chemical cost varies from $0.0011 to $0.0116 per barrel. For low-pressure cracked gasoline the chemical cost runs from $0.0017 to $0.0109 per barrel, and on the high pressure cracked gasoline from $0.0017 to $0.0124 per barrel. The range of volumetric loss in the process is from 0.1 to 0.75 percent and the average is very close to 0.25 percent. Degradation of quality has not been sufficiently accentuated as regards sweetening until recently. It is now apparent that as much as 3 octanes may be lost with an increased tetraethyl lead consumption of .4 cc. per gallon of gasoline. The value of precaustic washing in the sweetening process to effect the complete removal of mercaptans and thus minimize the octane loss and increased consumption of tetraethyl lead is discussed by Dr. Rippie.[5]

If gasoline is caustic washed rather than sweetened, it is estimated that a savings in chemical cost of from 70 to 90 percent and a reduction of around 50 percent in processing cost would result. And, since there is no degradation of octane value, the increased tetraethyl lead requirement noted above is automatically eliminated. Another important consideration is the effect of excess sulfur left in gasoline by doctor sweetening on increasing anti-oxidant requirements. However, since under proper conditions as described a satisfactory product can be made by the proper selection or combination of these processes followed by washing with caustic soda solution, the authors find it difficult to justify doctor sweetening by any process. Considering the almost complete elimination of contact of the motorist with the odor of the gasoline he is using and the debatable nature of this quality, it hardly makes a reasonable or substantial standard on which to base a refining process.

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Design of Oil Absorbers

Factors Governing Their Performance

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THE oil absorber is one of the most important units making up a natural gasoline plant. Failure on the part of the absorber to remove a high percentage of the desirable constituents from the raw gas stream may cause the plant as a whole to operate at a loss no matter how efficiently the distillation and fractionating units may perform. It is paramount, therefore, that every employe who is charged with the control of plant operations such as the chemist, gas tester, stillman, and superintendent, have an understanding of the principles governing the absorption process[8 12].

In every commercial process involving either a physical or a chemical reaction the two foremost factors to consider are:

1. What are the equilibrium conditions between the reacting substances which will prevail under different temperatures and pressures?

2. At what rate will the reactants approach equilibrium?

THE EQUILIBRIUM FACTOR

Equilibrium requires consideration although its conditions are unobtainable in an actual process; in fact a close approach to equilibrium should not be attempted unless a good price for gasoline justifies a high degree of extraction efficiency.[7] Without a well denuded oil, quite low in vapor pressure, an absorber must of necessity release some gasoline vapors into the residue gas even should equilibrium conditions exist. For example, an absorber operating under a pressure of 50 pounds per square inch absolute and 90°F. at the top is supplied with a lean oil having a vapor pressure of three millimeters of mercury or 0.058 lbs. per sq. in. absolute at 90°F. If it is assumed that the vapor pressure is substantially all due to the presence of normal-pentane (N C_5H_{12}) in the oil, the percentage by volume (or mol-per cent) in the residue gas may be easily calculated from Raoult's Law. This law states that the partial pressure of any constituent in the gas phase is equal to the partial pressure of the same constituent in the liquid phase, or mathematically

Partial Pres. in Gas πy = Partial Pres. in the liquid px (1)

Where π = Total Pressure (50 lbs. in this case)
y = mol fraction in the gas (mol. frac. of N C_5H_{12} in this case)
p = vapor pressure of the pure component (or N C_5H_{12})
x = mol. fraction in the liquid (or N C_5H_{12} in the lean oil).

Substituting in equation (1)

$50\ y_N\ C_5H_{12} = .058$
$y_N\ C_5H_{12} = .0016$ or 0.16 per cent by volume in the dry residue gas under equilibrium conditions.

In 1000 cu. ft. of residue gas there would be 1.6 standard cubic feet of normal pentane (.0578 gallons) or about .082 gallons of (26-70) natural gasoline. Under actual plant operation a residue content of 0.1 gallons of (26-70) natural gasoline might be obtained in this case.

For higher pressures Raoult's Law does not hold[2 4 6] since the hydrocarbons do not exert their true vapor pressures. For such cases actual experimental data based on Henry's Law is recommended. This relationship, giving the distribution of each constituent between liquid and vapor has been expressed by Brown, Souders, et al[2] as:

$$y = K x$$

Where y = mol. fraction of one constituent in the vapor
x = mol. fraction of the same constituent in the liquid.
K = a partition coefficient based on temperature, pressure, and percentage of solute in solvent.

The lower the percentage of any constituent in any solution the greater its actual vapor pressure or fugacity.

THE RATE OF ABSORPTION

The speed at which any process will proceed varies directly as the driving force or potential, and inversely as the resistance against which the reaction must go. The driving force in the case of oil absorption depends upon the difference in partial pressure between the gasoline constituent in the gas and the partial pressure of the same constituent in the oil. This force is generally great at the base of an absorber, but falls off rapidly as the top of the absorber is approached. Walker, Lewis and McAdams[12] discuss at length Whitman's theory whch states that the resistance opposing the absorption is caused by stagnant films of gas and oil adjacent to the interface between the two phases. If the gas is methane, which is relatively insoluble in oil, the resistance is controlled largely by the liquid film, but with normal pentane which is highly soluble in the oil, once it comes into contact with the oil the controlling resistance is the gas film. At the interface between the two phases equilibrium conditions are assumed to exist.

THE TUG OF WAR BETWEEN METHANE AND THE ABSORPTION OIL

Oil absorbers are sometimes built as packed columns but more often as bubble-plate towers in which the gas stream rises counter-currently against the descending absorption oil. Within the tower the reaction may be compared to that of a "tug of war"

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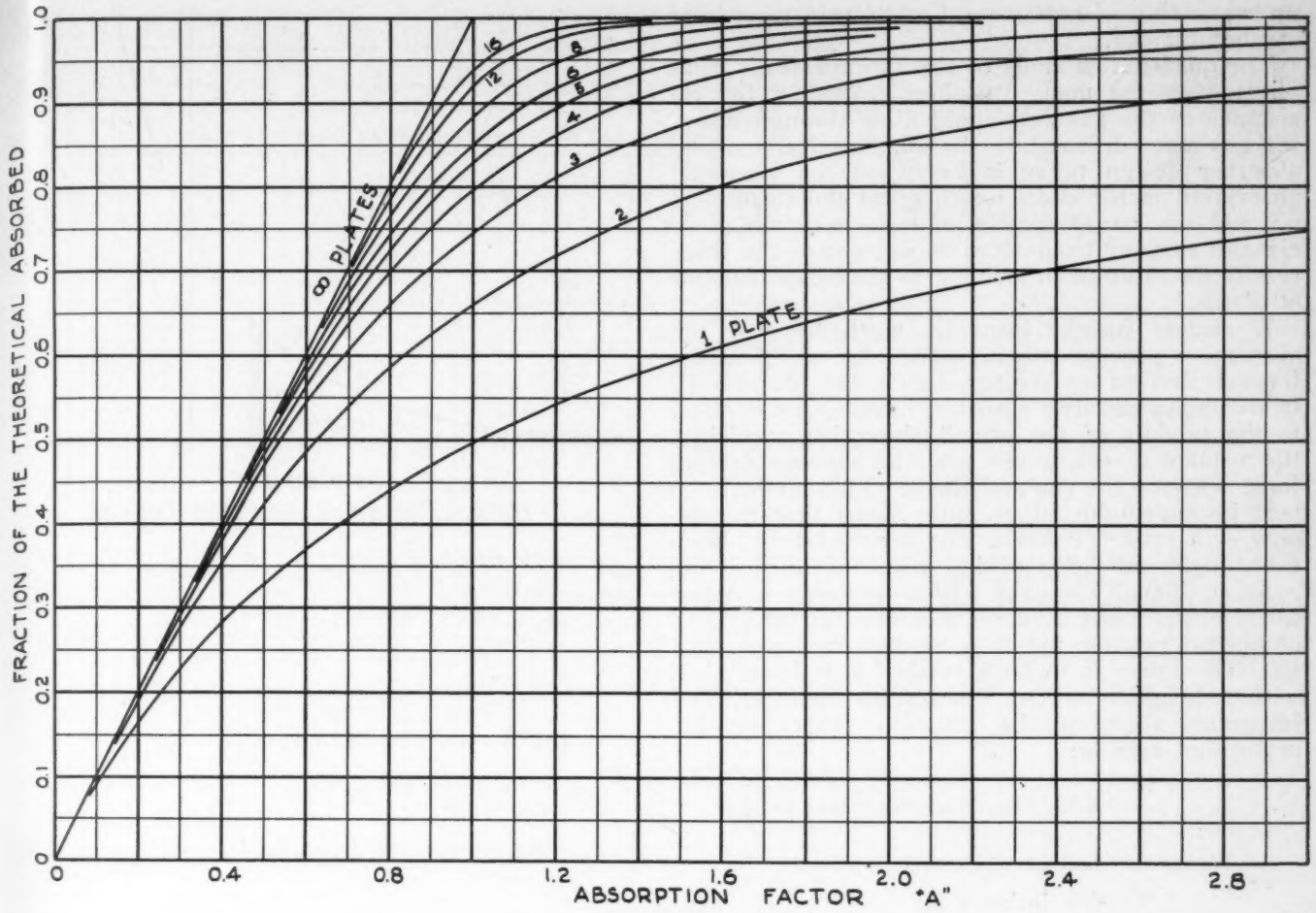

FIGURE 1

Kremser Absorption-Factor Chart.

between the methane (usually the predominating component in the dry gas) and the lean absorption oil over the possession of pentane or other desirable fractions. At first thought one would say that a rich gas should require more absorption oil than a lean gas to bring about a certain degree of extraction, but a little study will show that the leaner gas requires more oil. It may be explained qualitatively by bringing out the fact that the drier gas has more methane in it to pull the pentane over into the residue gas. The following example is given for a quantitative proof.

Two gases having 5 percent and 20 percent by volume of normal pentane respectively are to be absorbed in separate towers each having 12 theoretical trays and the same pressure and temperature conditions. The dry gas is to contain only 0.2 percent by volume of N-pentane. What will be the relative oil circulation for each tower based on 1000 standard cubic feet of raw gas, assuming the lean oil to be free of all gasoline constituents?

Solution:

By material balances it may be shown that the gas containing 5 percent N-pentane has about 96 percent of its pentane absorbed, while the 20 percent gas has approximately 99 percent absorbed. Referring to Figure 1, the absorption factor[2 5] for the 5 percent gas will be 1.155, and for the 20 percent gas, 1.32. These absorption factors are proportional to the lean oil circulation per thousand cubic feet of lean gas, since the factor is defined as

$$\frac{L}{VK}$$

where L = mols of lean oil
V = mols of lean gas
K = Equilibrium or partition constant.

For the 5 percent gas the ratio of mols of lean oil to mols of rich gas will be 1.155 x .952 or 1.10, since there are 952 cubic feet of residue gas for every 1000 cubic feet of raw gas. With the 20 percent gas the same oil-rich gas ratio will be 1.32 x .802 or 1.058, since there are 802 cubic feet of residue gas per 1000 cubic feet of raw gas. The lean-oil circulation for the rich gas will therefore be $\frac{1.058}{1.10}$ or .96, or 96 percent of the amount required for the lean gas. In actual operation, however, the ratio is still more favorable for the rich gas since the fugacity of the pentane goes down when larger percentages of pentane are dissolved in the absorption oil.

ABSORBER CAPACITY

The size of an absorber is governed largely by two factors:

1. Entrainment.
2. Rate of solution of the gas into the oil.

Entrainment is the mechanical carrying of liquid droplets upward from one tray to another. Various factors affecting entrainment such as tray spacing, vapor velocity, liquid rate, and down-spout capacity. have been studied[1 3 9 10 11]. Further experimental work on the priming and flooding characteristics of bubble-plate towers needs to be done before a better

understanding of entrainment and liquid carry-over can be obtained.

The quantitative study of absorption rates is complicated by the many variables governing the resistance of the gas and liquid films through which the gas must diffuse into the oil. Practically all oil absorber design today is based on the Kremser[5] absorption factor chart which gives the theoretical but not the actual number of trays required. Efficiencies ranging from 50 to 80 percent of the theoretical are assumed in arriving at the actual number of plates.

A clearer insight into the mechanism of the absorption process can be gained by applying the formula derived by Walker, Lewis and McAdams[12] in which the quantity absorbed in unit time is equal to the product of the overall absorption coefficient, the volume of the tower, and the average driving force between the gas and the oil. This method can best be shown by an example illustrating the design of a packed column. The same analysis, however, might well be applied to a bubble-tray column.

An equipment company wishes to design a packed tower of sufficient capacity to absorb 10,000 gallons of normal pentane per 24-hour day from a natural gas. The tower is to be a vertical cylindrical shell with a height equal to 8 times its diameter. The integrated form of the equation expressing the absorption rate is:

$$\frac{G}{\theta} = KAL\ (P_{gas}-P_{oil})\ \text{log means average}$$

Where G = Gallons of pentane absorbed

θ = time in hours

A = cross sectional area of tower in square feet

L = height of tower in feet.

P_{gas} = Partial pressure of N-pentane in gas in pounds per square inch absolute

P_{oil} = Partial pressure of N-pentane in gas in pounds per square inch absolute

K = absorption coefficient in gallons per hour per cubic foot of free volume of the tower per log mean average partial pressure difference in pounds per square inch between the pentane in the gas and in the oil.

Data: Absorber pressure = 50 pounds at top and 52 pounds at base
Mol percent N-pentane in lean gas = 0.2 percent
Mol percent N-pentane in lean oil = 0.395 percent
Mol percent N-pentane in rich gas = 10 percent
Mol percent N-pentane in rich oil = 15 percent

Absorber temp. = 100°F. at top and 120°F. at the base

K = 1.15 = Absorption coefficient

Calculate diameter and height of tower. Assume packing in tower takes up a negligible amount of space. Solution:

Vapor Pressure N C_5H_{12} at 100°F. = 15.5 lbs. per sq. in. abs.
at 120°F. = 22.4 lbs. per sq. in. abs.

Driving force at top of tower:
$P_{gas} - P_{oil} = (50)\ (.002) - (15.5)\ (.00395)$
$= 0.1 - 0.0613 = 0.0387$ lbs. per sq. in.

Driving force at bottom of tower:
$P_{gas} - P_{oil} = (52)\ (0.1) - (22.4)\ (0.15)$
$= 5.20 - 3.36 = 1.84$ lbs. per sq. in.

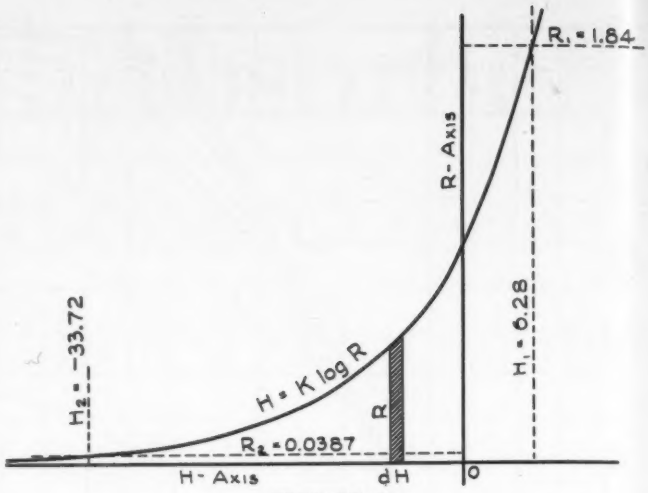

FIGURE 2
Graphical Analysis of Absorption Process.

Log mean driving force between the pentane in the gas and in the oil =

$$\frac{1.84 - .0387}{2.3\ \log_{10}\frac{1.84}{.0387}} = \frac{1.8013}{(2.3)\ (1.676)} = .4675 \text{ lbs. per sq. in.}$$

Let D = diameter of tower

$$A = \frac{D^2}{4} \text{ and } L = 8D$$

Then substituting into the formula $\frac{G}{\theta} = KAL\ (P_{gas} - P_{oil})$ log mean average

$$\frac{10000}{24} = (1.15)\ \frac{(D^2)}{4}\ (8D)\ (0.4675)$$

$$40000 = (24)\ (1.15)\ (3.14)\ (8)\ (0.4675)\ (D^3)$$
$$D^3 = 1232 \text{ or } D = 4.97$$

Take Diameter of tower = 5.0 ft.
Height of tower = (8) (5.0) = 40.0 ft.

QUANTITY OF GAS ABSORBED VS. TOWER HEIGHT

It is of some practical importance to know the quantity of gas which has gone into solution at various levels in the absorber. On rich gases some absorbers run as much as 20° to 30°F. higher at the base than at the top, necessitating the installation of internal cooling units within the absorber. By knowing where the bulk of the absorption takes place, these cooling units can be more accurately located in the tower. Such coolers rarely prove economical, however, in view of the small temperature difference between the rich oil and the cooling water.

The following calculations were made in order to determine the height in the tower (from the preceding problem) at which 50 percent of the gas had been absorbed.

The expression for log mean average between two numbers is equal to the integrated area under a simple log curve divided by the distance between the limits. Since the log mean average was taken between the driving force at the bottom of the tower, and the driving force at the top of the tower, the relation between tower height and driving force would be represented by the equation

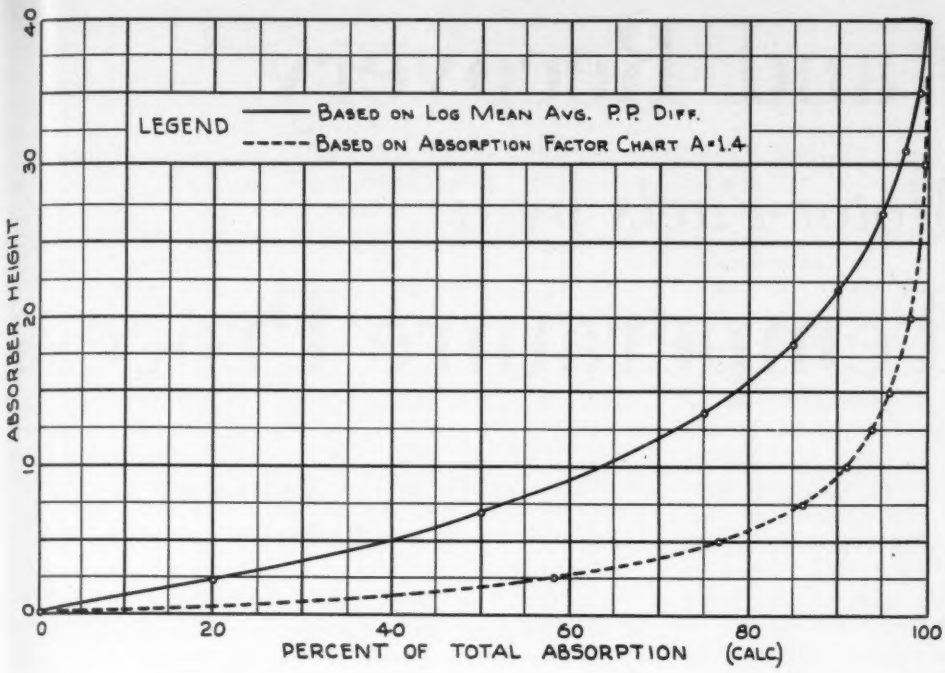

FIGURE 3

Percentage of Total Absorption at Various Levels Above Base of Absorber.

$H = K \log R$

Where $R = P$ = Driving force (See Figure 2)

When $H_1 = ?$ $R_1 = 1.84$

$H_2 = ?$ $R_2 = .0387$

$$H_1 = K \log R_1 \quad \text{or} \quad K = \frac{H_1}{\log R_1}$$

$$H_2 = K \log R_2 \quad \text{or} \quad K = \frac{H_2}{\log R_2}$$

$$\frac{H_1}{\log R_1} = \frac{H_2}{\log R_2} \quad \text{or} \quad \frac{H_1}{.2645} = \frac{H_2}{-1.421}$$

$$.2645\ H_2 = -1.421\ H_1$$

$$H_2 = -5.37\ H_1$$

$$H_1 - H_2 = 40 \quad \text{or} \quad H_2 = H_1 - 40$$

then $H_1 - 40 = -5.37\ H_1$

$-40 = -6.37\ H_1$ or $H_1 = 6.28$ (upper limit)

$-H_2 = 40 - 6.28 = 33.72$

$H_2 = -33.72$ (lower limit)

$6.28 = K\ (.2645)$ or $K = 23.7$

$-33.72 = K\ (-1.421)$ or $K = 23.7$

Log Mean Avg. 0.0387 and 1.84

$$\frac{1.84 - .0387}{2.3 \log_{10} \frac{1.84}{.0387}} = \frac{1.8013}{(2.3)\ (1.676)} = .4675$$

Area = (Log Mean Avg.) (Length) = (.4675) (40) = 18.7

½ Area = 9.35

Area under the log curve is representative of the amount absorbed. ½ Area would represent 50 percent absorbed.

$$\text{Area} = \int_{H_2}^{6.28} R\,dH \qquad H = K \log_{10} R \text{ and } R = 10^{\frac{H}{K}}$$

$$\text{Area} = \int_{H_2}^{6.28} 10^{\frac{H}{K}}\,dH$$

This integral is of the form

$$\int b^{au}\,du = \frac{b^{au}}{a \log_e b} + C$$

In this case $b = 10$

$$a = \frac{1}{K} = \frac{1}{23.7} = .0422$$

$u = H$

$$\text{Area} = \left[\frac{10^{.0422H}}{.0422 \log_e 10} \right]_{H_2}^{6.28} = 9.35$$

$$9.35 = \frac{10^{.265}}{.096} - \frac{10^{.0422H_2}}{.096}$$

$.898 = 1.84 - 10^{.0422H_2}$

$.942 = 10^{.0422H_2}$

$\log .942 = .0422H_2$

$-.026 = .0422H_2$ or $H_2 = -\frac{.026}{.0422}$

$H_2 = -.615$

$H_1 - H_2 = H$ = Height above bottom at which 50 percent of the gasoline would be absorbed

$6.28 - (-.62) = H = 6.9$ feet

Figure 3 is presented in order to show the percentage of the total amount of the gas absorbed at various levels in the tower. Although the curve based on the log mean average driving force falls somewhat above the one taken from the absorption factor chart, it is quite evident that in either case, by far the greater part of the absorption takes place in the first few trays near the base of the absorber. From this fact it is easy to see why the temperature rise is pronounced as the oil approaches the bottom of the absorber.

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Refining and Recovery

Of Light Motor-Fuels by a

Low-Temperature Process

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A DESCRIPTION is presented dealing with new processes for the recovery and refining of low-boiling motor fuels—both of which are based on low-temperature procedures.

The first type of refining deals with the recovery of white benzol from crude benzol by means of chilling and crystallization operations, followed by removal of impurities from the crystallization benzol by distillation.

Secondly, a procedure is described for recovery of light benzene or gasoline from mixed gases, obtaining as by-products substantial quantities of propane and butane.

MOTOR BENZOL

The materials processed are:

Crude benzol to white benzol to motor benzol.

In contrast to earlier processes, according to which pure benzol is said to have been obtained from crude benzol by means of chilling in a series of steps, which were found to be impractical for commercial operation, the process (1) to be described operates in one chilling-step, resulting in practically a complete removal of impurities of the benzol obtained by crystallization.

The procedure is based on the fact that one is able to use to advantage quite completely the cold or low temperature of crystallized benzol by use of suitable liquids, which efficiently exchange low temperatures with the crude benzol. Suitable liquids are: pure benzol; water; diluted alcohol or benzene.

The crystalline mass obtained by cooling the crude benzol is freed of mother liquor most completely by centrifuging. If the crystal mass is now warmed almost to the melting point of benzol, a small amount of wash liquid is obtained; with removal of the latter by centrifuging, a consequent removal of impurities is effected. The benzol is now gradually melted in the centrifuge with one of the liquids mentioned above which absorbs the cold from the crystals.

A practical example is given. One hundred kilos of crude benzol containing 60 per cent of C_6H_6 is pumped into a closed vessel equipped with stirrer and cooling coil. The coil is connected to an ammonia refrigeration system. When the mass reaches a temperature of 35°C., it is dumped through a short hopper to the centrifuge. The centrifuge liquid or mother liquor, high in toluol content, flows to a receiving tank. A small amount of 30 per cent diluted alcohol is now sprayed on the crystal-mass until the centrifuge liquid is only slightly colored and shows an m.p. of 0°C. The material in the centrifuge is now treated with 100-150 kilos of warm dilute alcohol. After a short period of time the crystal-benzol (about 50 kilos) is found to be melted. This liquid will show a melting point of practically pure benzol. It contains at most only small amounts of unsaturated hydrocarbons or sulphur compounds.

In order to make use of the fairly low temperature of this liquid, it is pumped to a heat exchanger, in which the crude benzol is pumped counter-current to the pure liquid. The crude benzol leaves this exchanger at 7°C., and enters a second exchanger, in which it encounters the cold mother liquor in counter-current fashion. Raw benzol leaves the second exchanger at —10°C., a temperature close to that at which solid benzol will separate; it then empties into the mixing vessel, in which approximately 80 per cent of the benzol separates in crystalline form. The pure, warm final product (from first exchanger) is separated from the chilling-solution (here, aqueous alcohol) in a settling chamber.

Instead of exchanging the low temperature of the separated product with or to the crude benzol, the low temperature can be used to supercool in part or completely the cooling medium (for example, liquid ammonia); by this means a decrease in the energy requirements of the refrigeration system is secured.

As liquids to supply the cold, glycol, glycerin and similar liquids can also be used besides benzol, water or aqueous alcohol.

The adoption of this new process is especially profitable in connection with recovery of benzol from gases by low temperature application. In such cases the crude starting material is not contaminated with constituents of the "wash oil," and its benzol content is higher than "wash oil" benzol; for this reason the crystallization process becomes more practical. By a combination of both procedures, a common source of refrigeration can be employed, securing low operating costs, and a simplification of operations.

It has been found, however, that even traces of impurities in the crystallization benzol thus obtained affects the qualities of the product in storage, resulting in gum formation. Besides, the by-product mother liquor, containing nearly all the impurities, is not commercial as it stands. Sulphuric-acid refining of both the pure benzol and mother liquor is of no practical value. For this reason, the originator

of the process described above, Dr. Schuftan, of Linde's Ice Machine Company, of Hollriegels-Kreuth, Munich, Germany, proceeds as follows: it consists in a rectification of the crystal benzol in a special manner,[2] and comprises a purely "physical procedure"—based on the fact that rectification separates from both benzol and mother liquor certain impurities, the result being products far superior to those commercially available and made by the usual refining procedure.

The mother liquor, a by-product of the crystallization process, is distilled and a cut up to 78°C. is separated; this fraction contains the greatest portion of unsaturated hydrobenzoles, which distill off in azeotropic mixtures. These hydrocarbons are easily resinified and are generally in ordinary processes very difficult to remove, since their boiling points are very close to that of benzol. Since they are fairly well concentrated in the mother liquor, they distill from the latter in azeotropic mixtures.

The next fraction from the mother liquor to 92°C. contains nearly all the benzol, although still traces of impurities such as unsaturated hydrobenzoles. Such impurities are removed by adding this fraction to the crude benzol, to be crystallized as described above (main patent). Even thiophen is removed from the crystallized pure benzol in this manner.

The primary crystalline benzol is also freed from last traces of impurities by distilling off 2.5-4.5 per cent of a primary or first cut. By rectification one obtains a decrease of unsaturateds and sulphur compound with results as follows:

Benzol	Crystalline Benzol 99%	Crystalline Benzol After Rectification 99%
Melting Point	5.3°C.	5.52°C.
Bromine Value	0.7	0.1
Sulphur	0.08 per cent	0.2 per cent

The explanation for this improvement by distillation is that the impurities are not so much derived from the adhering mother liquor but that they form mixed crystals composed of furan and pyrrol compounds.

According to the new process, it is possible to produce motor benzol in good yield and of high purity from crude benzol by means of crystallization and distillation—a procedure of purely physical means.

The impure by-products from crude benzol purification, which are completely destroyed in the usual sulphuric-acid refining processes can be used as solvents, or where they are derived from gas-house benzol, they may be used to enrich or increase the B.t.u. value of illuminating gas.

The by-products, comprising 20-27 per cent of crude benzol and containing 95 per cent of the unsaturateds and 80 per cent of the sulphur compounds originally present in crude benzol, can be easily converted to low-boiling motor fuels by destructive hydrogenation.

The benzoles and toluols made by this new process are adaptable for conversion to high grade commercial products after a light treatment of the vapors with calcium permanganate or perborates.

II—LIGHT BENZINE FROM MIXED GASES

Light benzene is normally obtained from natural gas, cracked or coking-still gases by compression and low temperature absorption. The product so obtained, which is liquid, contains considerable very volatile constituents such as propane or butane, which must be removed by "stabilization" before use. To remove these volatile fractions, stabilizers heated by steam usually are employed.

A process developed by Dr. Schuftan[3] of the Linde Ice Machine Company, A.G., combines in a simple and practical way the condensation of light benzene with the stabilization. The volatile or gaseous constituents are driven off by the aid of circulating a low-temperature carrier or cooling agent in heat-exchange conditions with the raw gases. The low temperature produced by evaporation of the most volatile portions from the primary light benzene is made use of to condense the light benzenes from the raw incoming gas. Liquid low-boiling benzene is used as a carrier-exchange medium for the low-temperature evaporation exchange. The lightest or most volatile portions transfer their "cold of evaporation" to condense or super-cool a refrigerating liquid such as ammonia, through whose evaporation the crude gas is again cooled; by these means the normal constituents of light benzenes are obtained in high yield, besides large quantities of lower-boiling materials, such as propane and butane.

The sketch (Figure 1) describes the procedure encountered.

Compressor 1 takes suction from a supply of gas —preferably purified beforehand, and compresses it to a critical pressure of 3-4 atmosphere (liquefaction). The heat of compression is removed by water-cooler 2. The gas mixture is in counter current flow in alterate coolers 3 (a) or 3 (b) with gas, which has been freed of its benzene constituents, and the condensate from this cooling step is collected in receiver 4(a). One of the coolers 3(a) or 3(b) acts in a manner known to the art as a gas cooler, while the other one is being warmed in order to melt the solid condensate separated in it—especially ice. Further cooling of the gas containing light benzenes takes place in cooler 5—in heat exchange with ammonia evaporating at reduced pressure; by this means a cooling temperature of 50°C. is reached. The condensates separating at this temperature are caught in separator 4(b). The remaining benzene which the gas-mix still contains at 50°C. is condensed by expansion of the gas by supplying external work thereto. The expansion takes place in expansion machine 6, which is invariably coupled with compressor 1, using piston machine. A turbine can also be used, coupled to a turbo-compressor. A temperature of 80°C. can be attained by expansion of the gas, at which point practically all valuable benzene-type gas constituents are now condensed. The condensates formed by this expansion are caught in receiver 4(c).

The stabilization of the separated light benzene in 4(a), 4(b) and 4(c) takes place in stabilizing column 7 into which the condensates are introduced at proper heights according to their composition and in which propane, butane and true gaseous hydrocarbons can be separated.

The heat available for vaporizing the volatile condensate should in the opinion of the inventor be derived and obtained from the incoming raw gas. If one desired to heat column 7 by the raw gas the difficulty would follow that the heating pipes would soon be covered by condensed ice, etc. In view of this, one is forced for continuous operation to employ 2 rectifier columns which, as in the case of coolers 3(a) and 3(b), serving alternately for the

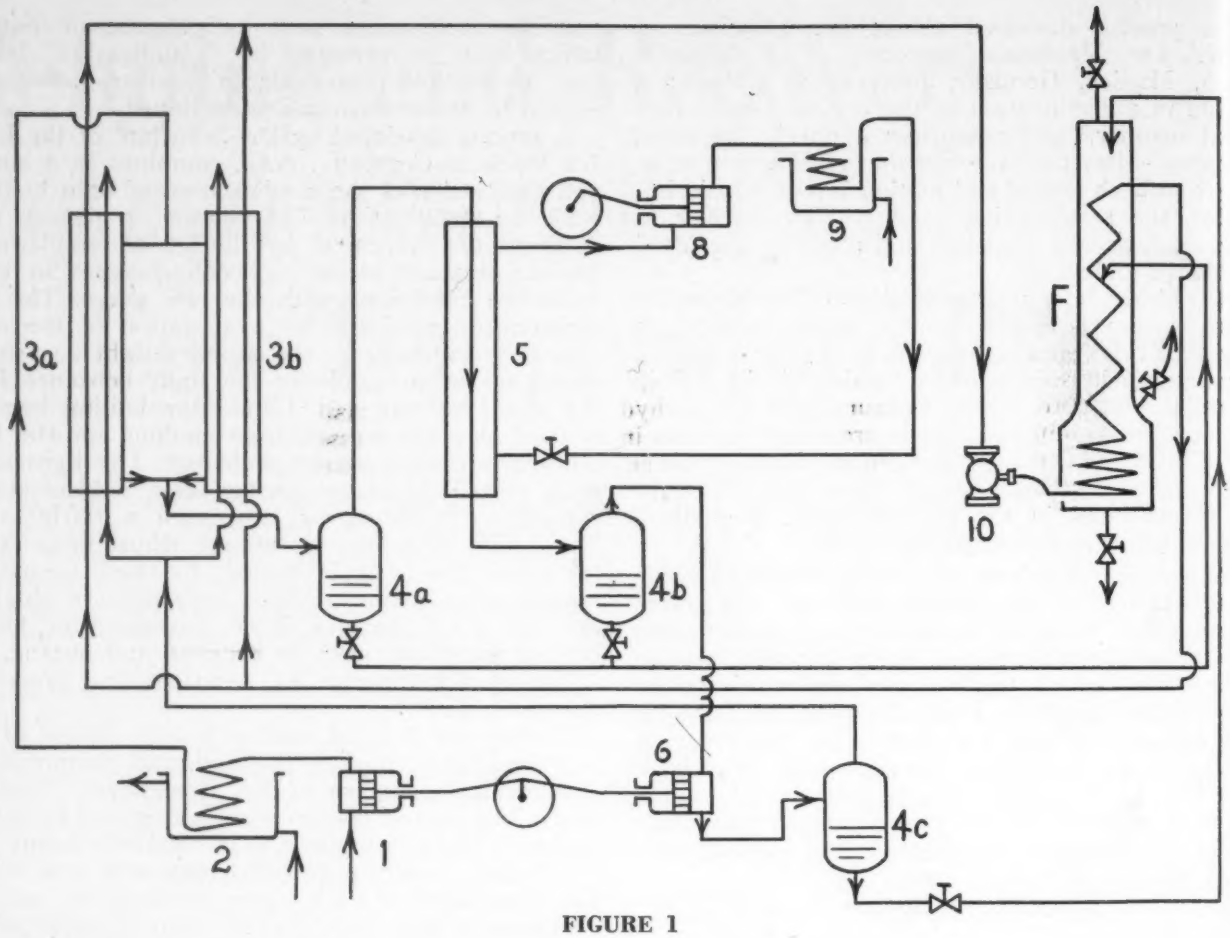


FIGURE 1

cooling of the raw gas, could be warmed and thawed out. Nevertheless one can operate with one column when the heat-exchange between raw gas and raw light benzene is carried out in a preliminary manner. In the sketch shown herewith the low temperature existing in the rectifier is applied to the raw gas by means of an exchange liquid or cooling medium, such as, for example, by salt solutions, or more practical, by the recovered finished light benzene itself. For this purpose then the cooling medium is carried counter-current to the raw gas in the coolers 3 (a) and 3 (b), and thereby warmed up, and then through liberation of heat in column 7 again cooled. The cycle of the cooling medium between column 7 and the coolers 3 is controlled by pump 10.

When the raw gases contain a high percentage of low-boiling benzenes, heavy condensation takes place in the expansion machines during the expansion of the gas; this causes improper operation of these expander pumps. In this case the gas leaving the separator 4(b) is not led into the expansion machine directly, but is "heat-exchanged" instead with the cold expanded gas coming from the expansion machine, before the gas enters the cooler.

Since the degree of refrigeration and the returns of energy-input in the expansion machine decrease with a lowering of the initial temperature of the entering raw material, the raw gas in this process is again returned to the cooler in which it comes in contact by heat-exchange with the expanded gas so that it only undergoes an intermediate cooling, entering the expansion machine at practically the same temperature, similarly, as if it had been led directly to the machine after the ammonia cooler. For this reason the gas is quite unsaturated since during the intermediate cooling a further quantity of light benzene has condensed, which materially taxes the expansion machine.

If a gas is to be processed which furnishes below 50°C. (behind the ammonia cooler) only small amounts of motor fuel condensate, then the expansion of the gas is no longer necessary for purposes of procuring low temperatures, since an ammonia compressor to produce cold then becomes much more practical than by using the expansion of the gas. In such cases a high energy-return can be aimed at with the expansion of the gases. For this purpose the gas is warmed up (before expansion) in counter-current heat-exchange with the raw gas. If one desires further to use up all the heat of compression, a temperature rise of over 100°C. can be attained. Expansion at this temperature results in a much higher energy-return in the expansion machine than expansion taking place at low temperatures.

Since the butane fraction is obtained by this process completely free of low boiling portions, it is possible to allow it to remain completely or almost entirely in the light benzene instead of separating it with the propane.

By this process it can be seen that yields of commercial liquid motor fuels can be obtained fully 33 per cent greater than those obtained by the usual compression and absorption methods.

[1] D.R.P. 622,727 of 5/14/1931.
[2] Addition D.R.P. 623,337 of Jan. 23, 1932 (Dr. Schuftan of Linde's Ice Machine Company, A.G.).
[3] D.R.P. 637,330 V. 10/11/1933.

Convenient

Constant-Temperature Bath for the Asphalt Laboratory

HAROLD BOTTOMLEY

DUE to the exact temperature limits necessary in asphalt penetration and ductility tests, a constant-temperature bath is a necessity in the asphalt laboratory. After seeing several in operation, and not being entirely satisfied with the results of any one; the one to be described was designed and assembled according to the following description.

The bath itself consists of a double-walled copper tub, the dimensions of which are given in Figure 1. The space between the two walls is filled with rock wool or other efficient insulator. One-inch copper tubes are soldered into the tub as shown, the one in the bottom being the discharge and the tangential one in the wall, the inlet. Circulation is maintained by a rotary or centrifugal pump placed under the bath. A medium sensitivity thermo-regulator is used to turn the 500-watt knife-type heater and the circulating pump motor on and off as needed. A resonant circuit-type relay is used as the switch for the heater and motor circuit. With this type of relay which must be mounted according to instructions with the instrument, the circuit is closed when the thermo-regulator contacts are open. When the bath comes to the predetermined temperature as set by the regulator, the contacts in the regulator close and the holding coil of the relay is shorted out; that is, the resonant circuit is disrupted and the relay contacts open.

As seen from the diagram, the motor starts and stops as heat is applied to the bath. At first glance this might appear to be a bad practice but a little reasoning will show otherwise.

The asphalt samples are placed on a false bottom about 1½ inches above the actual tub bottom. It is important that this zone where the samples are located be kept at 77° F. plus or minus .2°. The actuating bulb of the thermo-regulator is therefore placed in this zone. Even though the water in the bath is slightly cooler or warmer at the surface than in this region, the temperature at which the samples are kept will be correct. When the temperature falls sufficiently the heater and pump are automatically turned on and the contents of the bath are circulated several times before the pump stops. A pump should be used that will circulate from 25 to 30 gallons of water per minute.

A rotary pump was used in the original installation but since then its operation has called attention to the fact that a centrifugal would be more satisfactory. An ideal motor to use in this service is a one quarter-horsepower repulsion-induction type such as is used on household electric refrigerators.

The above operation has been based on the premise that the room temperature is lower than the bath temperature and that only heating is necessary to maintain the desired temperature. In summer when room temperatures run from 85 to 100° F. means must be provided whereby the bath can be cooled to 77° F.

Rather than attempt to thermostatically cool the bath to 77° F. it is preferable to cool at a constant rate and then control the temperature by thermostatically heating the water to 77°. This may be most conveniently done by placing a cooling coil in the bath and expanding into it ammonia or a propane-butane mixture. In most refineries that operate vapor-recovery systems, the latter method is probably more feasible and economical.

In building such a thermostat only first quality substantial parts should be used if full satisfaction is to be realized. The original thermostat described above has been in use for over two months and has not been touched for adjustment in that time. It has consistently maintained 77° plus or minus .1° F. over the entire period.

It is urged that the entire water containing system be constructed of copper or brass in order to eliminate rust formation; necessitating frequent changing of the bath water.

Figure 2 gives the electrical circuit in entirety, including the internal connections of the resonant relay. The bath is supported on a steel table of all welded construction, the pump and motor being on a shelf below. The heater and thermo-regulator are supported by means of a transite board shelf placed across a portion of the top edge of the bath. The switch box, fuse block and relay are mounted on a steel panel just below the edge of the table in easy reach of the observer.

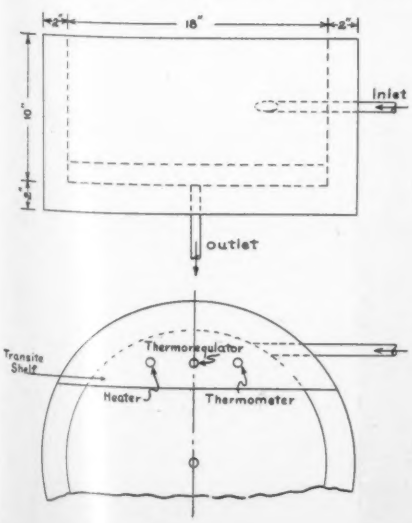


FIGURE 1

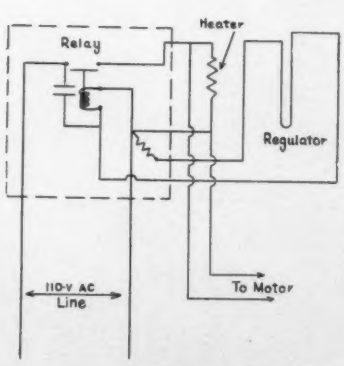


FIGURE 2

Time and Labor Saved

In Modern Compounding Plant

RECENTLY Freedom Oil Works Company, Freedom, Pennsylvania, built a completely new lubricating oil plant, in which various grades of motor oils, greases and heavy lubricants are blended from the base oils coming from the refinery. The building in which this plant is housed has an overall length of 243 feet and is 75 feet wide.

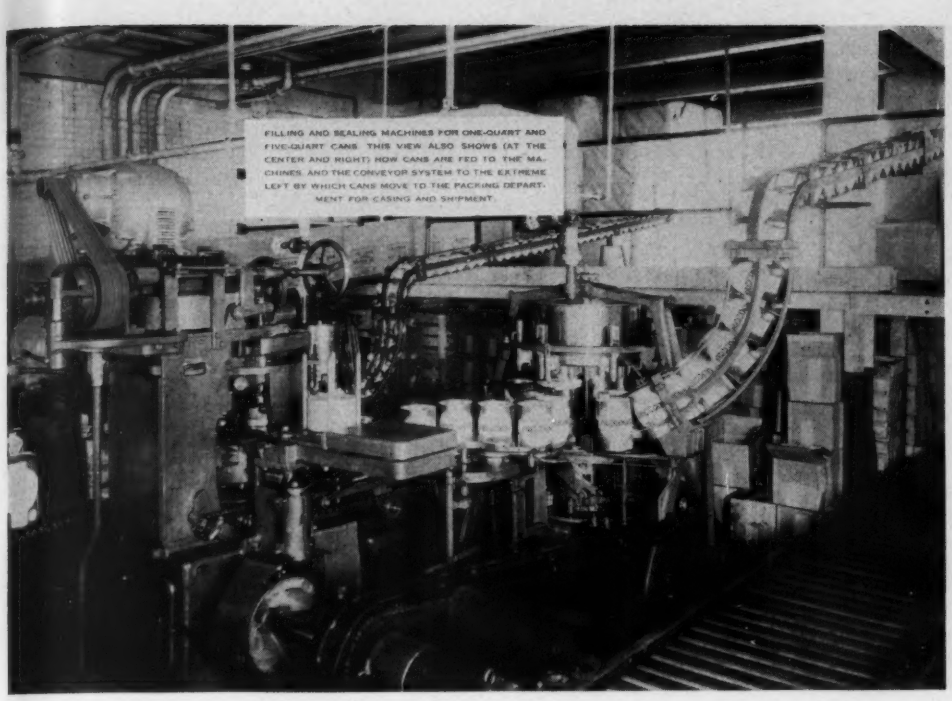
It was so designed that drums and all other packages, coming into the plant empty are unloaded from either trucks or box cars upon traveling conveyors that lead to any part of the building, upstairs or down, without the necessity of men lifting the cases. Likewise, when the containers have been filled, they move on other conveyors to storage, or back to the cars for shipment. By the arrangement of the various conveying systems, approximately twice the number of containers may be handled with half the number of men used before this plant was built.

New packages, cans, pails or drums, move directly to the package storeroom by the conveyor system, or, to the filling equipment, but old drums, coming back to the plant for refill, are sent to the reconditioning plant where all dents are removed, the paint cleaned from the exterior and all foreign material washed from the inside. This phase of drum handling, like all others in the plant, is "straight-line" operation, with the dented and dirty containers starting in at one door and continuing in almost a straight line until they are ready for refilling.

The first period of drum reconditioning is the preliminary inspection where those unfit for reuse are discarded, while those that may be placed back in service are placed on the straightening machine where hydraulic pressure, plus judicious tapping with a hammer, removes indentations. Bungs are then inserted in both openings after which the containers are subjected to a chemical bath which loosens the paint while mechanical arms equipped with stiff brushes thoroughly scrub the sides and ends. The solution used is caustic. Operators of the drum cleaning equipment do not handle the containers with their hands except to place them upon the first part of the machinery. Mechanical arms lift the drums from one operation to another throughout the cleaning process.

After the drums are cleaned of exterior paint they are thoroughly rinsed before being opened for interior cleaning. The drums are tilted over nipples through which flow a detergent which removes all traces of former filling. Hot water is used to remove the cleaning agent after which the drums are lifted to a slowly moving conveyor which runs through a gas fired oven to remove all traces of moisture. Another conveyor moves the drums to the painting

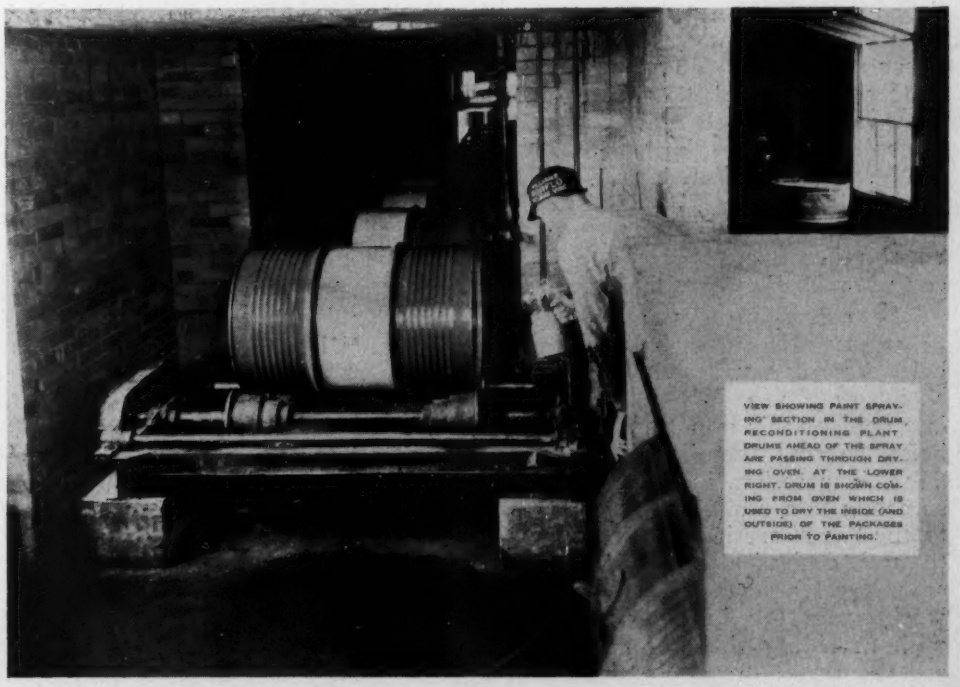
DRUM-FILLING PLATFORM, SHOWING ELABORATE CONVEYOR ARRANGEMENT. EMPTY DRUMS COME DOWN SPIRAL CONVEYOR, SHOWN AT RIGHT REAR, AND ARE SHUNTED TO THE VARIOUS AUTOMATIC FILLERS SHOWN AT LEFT AND RIGHT, EN ROUTE TO CARS AND TRUCKS.



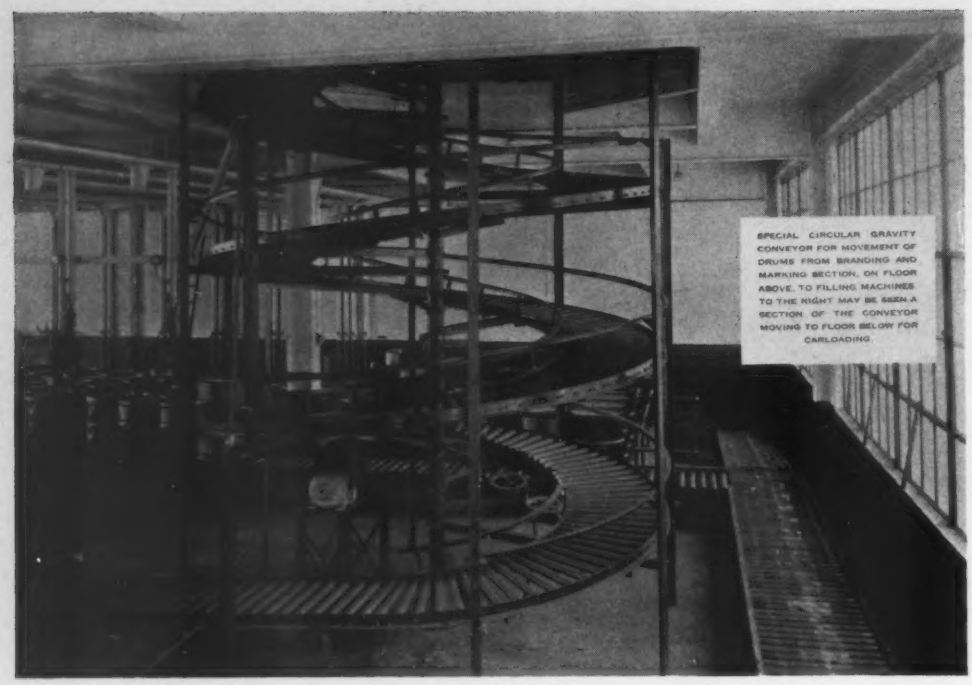
FILLING AND SEALING MACHINES FOR ONE-QUART AND FIVE-QUART CANS. THIS VIEW ALSO SHOWS (AT THE CENTER AND RIGHT) HOW CANS ARE FED TO THE MACHINES, AND THE CONVEYOR SYSTEM TO THE EXTREME LEFT BY WHICH CANS MOVE TO THE PACKING DEPARTMENT FOR CASING AND SHIPMENT.

hood. The painter handling this phase of the operation must place the drums upon the revolving wheels so that his spray gun can apply the colors evenly and without excessive covering. The travel of the drums through the various parts of the exterior, and interior cleaning, as well as subsequent travel after being painted, is synchronized to a speed that will permit each employee to do his bit without overcrowding.

As the drums emerge slowly from the drying oven, the painter swings them around and places each one upon the apparatus which revolves slowly, where he applies the distinctive colors to the heads and sides. After an even coat has been applied a button is pushed which sets in motion mechanical lifting arms which remove the painted drum from the rolls to the drying, or curing, oven conveyor. The curing oven is sufficiently long that a freshly painted drum is given sufficient time that when it arrives at the other end the paint is sufficiently hard that it will not peel. One drum follows another through this oven and when emerging at the opposite end, drops from the conveyor upon an upending machine, which changes position from reclining to standing. The apparatus between the curing oven conveyor and the long unit leading to the storage room was built with equalizing weights so arranged that when a drum drops into it, it swings over and around so that one end of the drum rests upon the long conveyor, which immediately removes the drum from the upending machine. As soon as it is freed of its drum, the automatic counterbalance causes it to assume a position against the curing oven so that the next drum may be handled. This operation con-



VIEW SHOWING PAINT SPRAYING SECTION IN THE DRUM RECONDITIONING PLANT. DRUMS AHEAD OF THE SPRAY ARE PASSING THROUGH DRYING OVEN. AT THE LOWER RIGHT, DRUM IS SHOWN COMING FROM OVEN WHICH IS USED TO DRY THE INSIDE (AND OUTSIDE) OF THE PACKAGES PRIOR TO PAINTING.


SPECIAL CIRCULAR GRAVITY CONVEYOR FOR MOVEMENT OF DRUMS FROM BRANDING AND MARKING SECTION, ON FLOOR ABOVE. TO FILLING MACHINES TO THE RIGHT MAY BE SEEN A SECTION OF THE CONVEYOR MOVING TO FLOOR BELOW FOR CARLOADING.

tinues with the regularity of clockwork as long as drums are fed into the beginning of the system at the initial cleaning equipment.

A thorough inspection is maintained throughout the plant so that defective drums are not permitted to pass into the building. Those unfit for carrying lubricating oils are stacked aside and used on company leases and around the refinery for various purposes. An inspection is carried on through the cleaning process, and if a drum appears to be serviceable before being put through the straightening, or dedenting process, and indicates by interior inspection by inserted lights that it will not carry lubricating oils, it is discarded at that point in the line.

Drums leaving the drying oven, or new ones coming from the receiving platform are directed to any part of the building by way of conveyors. If shipments demand that the drums be filled immediately they move to the automatic drum fillers by a spiral conveyor. To one not accustomed to a complete conveyor system, the filling machines seem to be completely surrounded by conveyors, radiating from a central track which moves the empty drums from the spiral conveyor. Drums move steadily down this track so that employees may direct their movements to the filler where oil is rapidly placed in the drum. By giving the filled drum a slight push, the conveyor takes it again and moves it to another long conveyor leading to either trucks or box cars.

The conveying system consists of innumerable rollers mounted upon ball bearings and actuated by belts directly beneath them. Rollers below, or at the side of the filling equipment, are "dead" in that there is a spot at this point which is not actuated by the belts, so that the drum is stationary while being filled. The rollers in the conveying system operate so easily that a slight push on a full drum moves it to the "live" section of the conveyor which takes it to any desired part of the building. Convenient cross-overs are placed at frequent intervals so men working in the building may move freely about without riding the conveyors. These conveniences are the result of observations carried on by A. J. Fruth, lubricating plant superintendent, and also the designer of the plant. Cross-bars are placed between several of the rolls at convenient points in the building which will support the weight of a man and prevent slipping upon moving equipment.

The plant is a complete oil blending, grease manufacturing, floor sweep preparing system, as well as producing some specialties such as vegetable oil soap. Raw oils for lubricating oil blending, compounding or mixing are received from the refinery nearby, in 25 storage tanks, each of which has a capacity of 22,500 gallons. The foundation for these tanks is in the basement, or ground floor of the building with heating coils placed beneath them, instead of within the tanks as is commonly done. By eliminating intimate contact between the oils within the tank and the steam coils, operations are simplified by eliminating many of the difficulties encountered in lubricating oil manufacture. Where bare steam coils are placed inside lubricating oil storage tanks, leaks develop after several months of service which causes cloudy oils. By boxing in the sills upon which the tanks are erected, very little loss of heat is encountered, and the time element is not excessive in bringing a tank of oil to the desired temperature for handling.

Rectangular tanks are installed in the lubricating oil blending department, each of which is 21 feet high, 11 feet deep and 9½ feet wide. These tanks were welded when fabricated and made in this shape to conserve space in the building. These tanks, like the raw oil storage tanks, are set on concrete sills for permanence, and arranged so that steam coils could be installed beneath them for warming or heating. Compressed air is used for creating sufficient turbulence in the blending tanks to completely mix the various ingredients used in making up various motor oils.

The air line inserted in the base of each tank crosses through the center and is equipped with jets three feet apart, which alternate in direction so that maximum stirring may be obtained.

Transmission of oils is obtained through three inch

pipe lines leading from motor driven centrifugal pumps on the first floor to the various tanks in the building. Remote control is used in this plant so that the operator desiring delivery of oil need only start the pump which is connected to a certain tank. Pipe lines are either overhead on supports attached to the building or placed in conduits in the floors. The conduits, or trenches, all lead toward a central sump so that all oil which might escape from packing glands, valve stem gates and the like, can be subsequently picked up, pumped back to the refinery, and rerun. So that friction could be eliminated as much as possible, pipe bends were utilized when that type of fitting fitted in with the general scheme of the blending plant.

Because of special oils sold to industrial institutions in the Pittsburgh area, where mixing could not readily be done with air for agitation, mechanical mixing kettles were installed equipped with apparatus similar to that used for making greases. Over these kettles, as well as those used for grease manufacture, hoods are placed which carry the odors and fumes from greases and special compounding materials through the roof and to the atmosphere. They are built with telescoping piping so that when a kettle was to be brought up to the necessary heat for mixing, the hoods may be drawn down against the cover. Counterweights are attached to these fume carriers so that they may be easily moved up or down.

One interesting feature connected with the lubricating plant, but not related to the petroleum industry, is the manufacture of soap. This production grew up like "Topsy" from a small batch made several years ago, until the output averages about 1500 pounds daily. It is manufactured from olive and cocoanut oils by a recipe worked up by a nationally known chemist. It is made in a kettle similar to grease making kettles, poured into wooden molds and cut after cooling into two pound cakes. This business has grown without especial pushing by the sales department, until the present output is not sufficient to fill current orders.

The conveyor system was placed throughout the entire building so that seldom, if ever, employees need lift a loaded container. The conveyor that was placed on the dock for car loading and unloading may be reversed in direction so that filled packages may be transferred to cars, or empty drums and cartons of cans may be placed in the building. It operates with so little effort upon the part of operators, that when a filled series of barrels are desired to be shipped in one car, with another series following, or mingled with the stream coming from the packaging room, a slight pull takes each one from the main conveyor to laterals extending into the box cars on the siding. Several cars may be loaded simultaneously with an employee standing beside the main conveyor to direct the individual barrel to the car in which it is to be placed.

Quart and five quart cans are fed to the automatic filling machines that are provided with conveyors on all essential sides so that as the containers are removed from the capping machines, they slide upon one of these conveyors which carries them to the carton filling department. After these cases have been filled, stenciled and sealed, they move to any part of the building desired. A feature of the conveyor system is shown by switches and right-angle turns in the tracks, adjustable and easily added to, so that it is extremely flexible. Five gallon cap sealed containers pass beneath the filling spouts upon conveyors so that a continuous, yet orderly, movement is maintained throughout this part, as well as other portions of the plant.

As the railroad occupies all the space on one side of the building, both ends blocked off from truck service, a loading recess was built into the front of the building that will accomodate three trucks at one time, all of which may be loaded simultaneously. Steel doors of the roll type were installed in this recess so that when not in use, it may be closed. Company trucks may also be stored in this space for the night, so that when an early start is desired for long hauls, the driver needs only to warm up the motor and drive out.

SECTION OF THE GREASE MANUFACTURING DEPARTMENT, SHOWING KETTLES SUSPENDED AT FLOOR LEVEL OVER WHICH ARE LARGE HOODS WHICH ARE LOWERED TO CARRY OFF THE FUMES. AT THE SIDE OF THE ROOM MAY BE SEEN STORAGE TANKS FOR STOCK OILS FOR GREASE MAKING.

Mud-Daubers *and* Motors

ELTON STERRETT

CONSEQUENCES of disruption of continuous processes in refining are usually serious enough to justify the installation of stand-by pumping units which, with the auxiliaries relied upon in case of fire, may be powered with electric motors, so connected as to be run from the refinery current sources as well as on purchased power.

These stand-by motors, especially where connected to purchased power lines only, in many refineries do not make a single revolution under power during an entire year, but require careful attention to insure their being in proper shape to assume the emergency load the instant it is switched to them. In one instance where the motors, utility powered with a high-demand charge, were regularly rotated by hand on the flexible coupling flanges at weekly intervals, the motors failed under load 10 minutes after being started under power after a 3-year period of idleness.

At first considered as being due to lack of or improper lubrication, examination of the least damaged of the rotor ends disclosed hard masses of oil-soaked clay still adhering to and blocking the passages between vanes of the built-in ventilating fan. Investigation of other stand-by motors in this same plant, also standing idle through considerable periods, identified these oil-soaked masses as the nests or egg cells of the common mud-dauber.

The dark interiors of the motors seemed to be especially attractive to the insects, the unbalanced load scraped from one end of the rotor in a 150 horsepower motor weighing 5.92 ounces; practically the entire amount being concentrated on less than 60° of arc around the fan or blower element. Assuming concentration at the center of this mass, and applying the formula

$$F = \frac{w \pi^2 N^2 r}{900 g}$$

where

w is weight in lbs. (5.92 oz. equals 0.37 lbs.)
N is the revolutions per minute or 3600
r is the radius of gyration, being 6″ or 0.5 feet
g is 32.2

The resultant force necessary to balance the centrifugal effect is found to be 816.5 pounds—ample to unbalance the high-speed rotor and force out or greatly reduce the effectiveness of the oil film which alone could oppose this distroting effect.

After thoroughly cleaning all stand-by motors, the vents through which the insects were passing were blocked with carefully fitted wooden plugs or stops. the edges being faced with felt to insure an unbroken and mud-dauber tight seal. After being fitted, these stops were painted a fire-extinguisher red, to make it readily apparent to any man starting the motors that the vents were closed.

As further precaution against starting the motors with these vent stops in place, a small panel of wood was made to fit over the handle of the main switch controlling these motors; this block also being painted red, and bearing on either side the warning

CLEAR
MOTOR
VENTS

so that, even if the color message of red stops failed to impress the man starting up the stand-by equipment, he must see the warning before actually switching the power to the motors so protected.

Tests of motors, with the stops intentionally left in place, showed that, apart from a stoppage again shortly after starting up, no material damage was sustained by the units. If run without ventilation, as is the case when the stops are left in place, the temperature-overload relay on the motor panels functions to cut the current before the internal temperature of the rotor windings is sufficient to weaken the insulation.

So effective have these mud-dauber stops proved in keeping out not only the insects with their concentrated lumps of mud, but also the drifting clouds of dust which from time to time sweep across East Texas from the "Dust Bowl" farther west, that one oil company has standardized on them for all motors not in daily service, providing sets for all motors in its possession, and issuing orders that they be used in case of shut-down or where intervals of more than 24 hours are to elapse between runs.

The upper right vent is closed with felt-edged stop, while within the vertical web of motor frame may be seen two of the mud-dauber accretions wrecking similar motor in stand-by service.

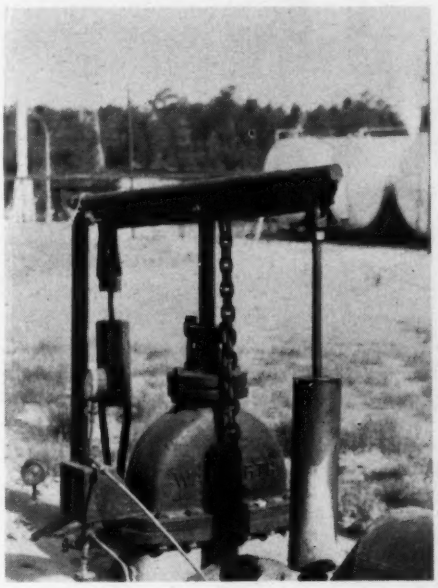
IDEAS to help plant efficiency . . .

Automatic Volume Controller

WITH two gas gathering systems coming into its plant from different sections of the East Texas field, and the amount of gas varying in volume abruptly, it was difficult for Gregg Tex Gasoline Company to obtain a constant suction on the compressors without loading them up at times. The system known as the West Gas System carries the larger load, and incidently, the leaner gas. To prevent abrupt surges from flooding the compressors beyond their capacity, an automatic valve was placed in the field line from the west system at a point where it was joined to the suction header at the plant.

A large valve having a rising stem, was flanged in the line at the point of connection to the header with the threads turned down so it could be operated as a quick opening valve. To operate this valve, a beam was placed above the stem, anchored at one end to a stationary post, and a piston rod attached to the other. A remote controller was placed on the body of the gate which admits pressure to the cylinder containing the piston attached to the free end of the beam, so that the gate is held open during normal operation. When a surge of gas enters the gathering line, reducing the vacuum held at this point pressure is automatically transferred to the opposite side of the piston which closes the valve to a point where the pressures are equal. The instrument on the body of the valve is easily adjusted to provide for any closing pressure and prevents an undue pressure building up on the suction header. J. C. A.

Tulsa, Oklahoma.



Automatic Volume Controller

A Muffler for Combustion Engine

IT IS necessary in some cases to muffle the exhaust and at the same time not cause a back pressure by obstructing the opening in the exhaust pipe. This is done by perforating the end of the exhaust pipe for about a foot and a half with a hack saw, putting as many cuts as possible in the pipe. Over these cuts wrap asbestos rope about an inch in thickness, cover this with a sheet of tin. In this way the exhaust gases are allowed to pass freely but the sound is muffled until the noise is not offensive. W.N.E.

Reducing Distillation Machine Operation Cost

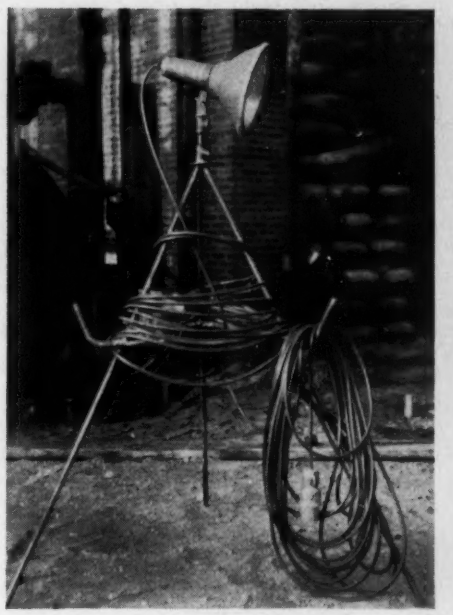
WHEN ammonia compressors are used in the laboratory to keep the bath temperatures of distillation machines within the range of 32° F. to 40° F. it usually is necessary to operate the compressor 24 hours per day. However, during the winter months outside temperatures often go below 40° F. for several months. During this time, by adding a coil, as shown in the drawing, to the circulating system the water baths can be maintained at the proper temperature level without the use of the compressor. Through the use of the by-pass the circulation can include the entire system. Then when the compressor is started no time is lost in cooling the brine tank which might otherwise reach room temperature. In the event a pour-point bath is maintained cool through the use of the compressor system, lower temperatures can be reached if the distillation baths are operated on a separate circuit. The reduction of operating costs pays for the installation of the additional coil in a short while.

M. E. WOOLDRIDGE.

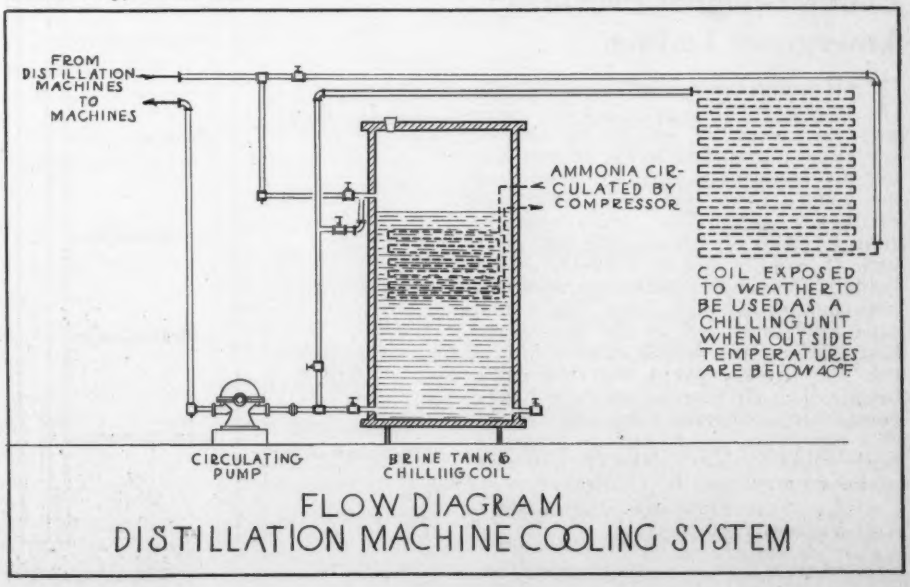
Ponca City, Oklahoma.

Portable Flood Light

WHEN working at night, or in portions of the plant where the light is not good, one naturally turns to a hand lamp for illumination. A hand lamp is not the most admired type of lighting for dark corners, because of the cord becoming entangled with tools, and when a helper holds the instrument so the rays will shine upon the darkest corner, the workman's head is invariable shutting off the light. Roosevelt Oil Company at Mt. Pleasant, Michigan, carries on repairs at night and also makes minor changes in



Portable Flood Light


FROM DISTILLATION MACHINES
TO MACHINES
AMMONIA CIRCULATED BY COMPRESSOR
COIL EXPOSED TO WEATHER TO BE USED AS A CHILLING UNIT WHEN OUTSIDE TEMPERATURES ARE BELOW 40°F
CIRCULATING PUMP
BRINE TANK & CHILLING COIL

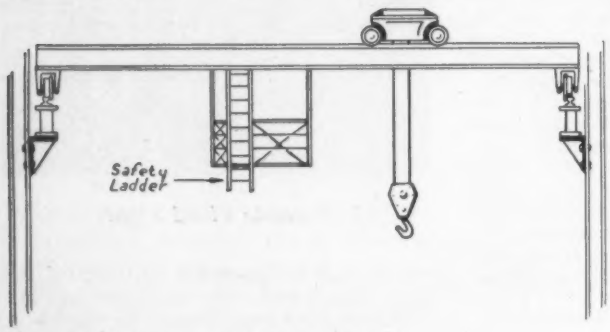
FLOW DIAGRAM
DISTILLATION MACHINE COOLING SYSTEM

the operation of the plant if it is needed. Plenty of light is necessary, and some one in the plant devised a portable flood light which may be carried by one man and adjusted so the rays will play upon a large area.

The tripod was made by using ⅜-inch pipe, welded at the upper ends to a flat disc, upon which was placed a nipple, also welded to the disc, but on the upper side. The nipple was flattened at the upper end so that a hole could be drilled in this part to receive a bolt which was fitted with a thumb nut. A small flood lamp was secured which had a strap surrounding the handle, so that when it was slipped over the flattened portion of the extended nipple, with the bolt in place, the rays might be directed at any desired angle.

To make the tripod more rigid, braces were welded from one leg to the other until it was surrounded by pipe holding the legs in the position most suitable to prevent tipping. The braces protrude beyond each leg, and were turned slightly upwards to contain the lamp cord when it was wound upon the tripod when not in use. The cord which attached to the lamp is also attached at the opposite end to an outlet connection so that more than one light can be operated off the one tripod, such as an extension hand lamp for inspecting dark corners where the rays of the flood light would not penetrate.

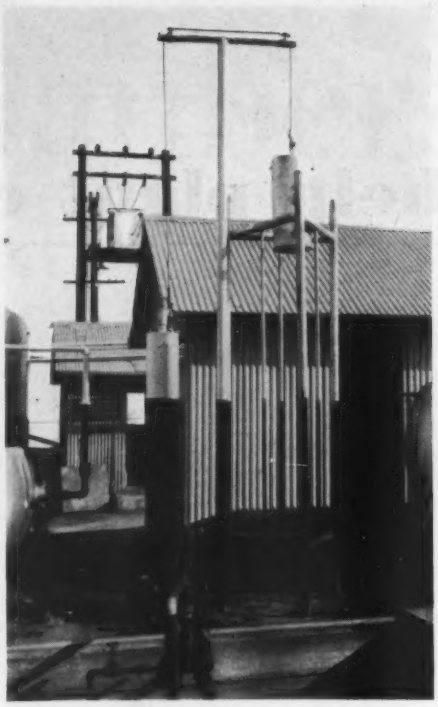
Both cords are equipped with flash proof attachments so that they can be used in the regular outlets around the plant, and each is provided with vapor-proof globes and guards. The men who use this lamp most say that it does the work very satisfactorily, but that it lacks a swivel at the upper end so that the lamp can be moved to direct the rays to another part of the ground, or vessel, without moving the entire tripod with its outlets and cords.



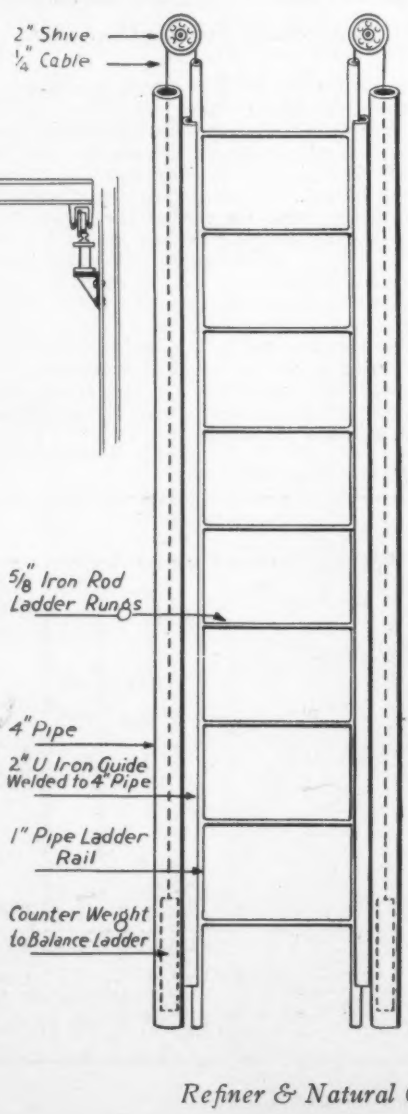
Counterweights Provided Emergency Ladder

THE counter-weighted ladder shown in the drawing provides for quick get-away from the cab of an electric traveling crane. The rungs of the ladder are made of ⅝-inch iron rods and are welded to the side rails, which are of 1-inch pipe. The counter-weight tube is made of 4-inch pipe with a 2-inch U-iron serving as a guide for the ladder. The cable is ¼-inch in diameter and runs on a 2-inch shive. The 4-inch pipe tube is closed at the bottom so that the counter-weight cannot fall to the floor in the event that the cable breaks. The counter-weights are heavy enough to compensate for the weight of a man, yet light enough to permit a fairly rapid descent to the floor in case of an emergency.

GEORGE B. MCCULLEN.
Gulf Refining Company,
Port Arthur, Texas.



After treating straightrun gasoline, the product is passed through a clay filter to remove all suspended particles. A counterweight is used to facilitate lifting of the filter shell when changing clay.


2" Shive
¼" Cable
⅝" Iron Rod Ladder Rungs
4" Pipe
2" U Iron Guide Welded to 4" Pipe
1" Pipe Ladder Rail
Counter Weight to Balance Ladder

Counterweight for Lifting Filter Shell

THE process used by Bachmann Petroleum Corporation in Signal Hill, California, for treating gasoline requires a filtration as the material leaves the last treating drum on its way to plant storage. Certain types of clays are used in relatively small quantities, and these require removal occasionally. The type of container used to retain the clay while the gasoline is passing through was made by using a joint of 15-inch pipe, flanged at the lower side and closed at the top. The base also supports an inner container which remains in place with the clay on the inside of it. To make changing easy, with little help necessary, the filter was mounted vertically. To further make replenishing clay easy, a counterbalance was attached to the filter case, which is suspended by a cable passing over pulleys inserted in the cross arm of a steel pipe post. A guard was built which encloses the travel of the counterweight, to prevent injury to workmen, should it fall due to breakage of the cable.

Improvised Hygrometer Easily Arranged

LACKING a ready made hygrometer, the following was constructed in about half an hour. A shelter was made of wood about two inches deep, or sufficient to protect the thermometers from the direct rays of the sun. The thermometers are supported in the shelter by means of corks. Slots are filed in the corks with a rat-tail file so that the thermometers just make a tight press fit. Two supports are used for each thermometer.

Improvised Hygrometer

An ordinary lamp wick is used to convey water to the bulb of the wet-bulb thermometer. This wick dips into a beaker of distilled water. The use of distilled water does away with scaling of the wick and thermometer bulb, necessitating cleaning very often if correct results are to be had.

HAROLD BOTTOMLEY.